The Power of Experience

Submitted To:

Simteche Hydrate CO_2 Capture Process

2006 Engineering Analysis Tasks of Contract Mod 017 - Final Report

Contract DE-AC26-99FT40248

Submitted By:

SIMTECHE

Los Alamos National Laboratory

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DISCLAIMER

Contents

Acronyms and Abbreviations

CO_2	Carbon dioxide
COE	Cost of Electricity
gmole	Gram mole
EPRI	Electric Power Research Institute
ETM	Engineering test module
KJ	Kilojoule
kW	Kilowatt
kWhr	Kilowatt hour
HDFR	Hydrate decomposition flash reactor
HFT	Hydrate formation temperature
IGCC	Integrated Gasification Combined Cycle
kWe	Kilowatt equivalent
kWh	kilowatt hour
LANL	Los Alamos National Laboratory
MW	Megawatt
NETL	National Energy Technology Laboratory
psia	Pounds per square inch absolute

As a result of an August 4, 2005 project review meeting held at Los Alamos National Laboratory (LANL) to assess the project's technical progress, Nexant/Simteche/LANL project team was asked to meet four targets related to the existing project efforts. The four targets were to be accomplished by the September 30, 2006. These four targets were:

- The CO_2 hydrate process needs to show, through engineering and sensitivity analysis, that it can achieve 90% CO_2 capture from the treated syngas stream, operating at 1000 psia. The cost should indicate the potential of achieving the Sequestration Program's cost target of less than 10% increase in the cost of electricity (COE) of the non-CO_2 removal IGCC plant or demonstrate a significant cost reduction from the Selexol process cost developed in the Phase II engineering analysis.
- The ability to meet the 20% cost share requirement for research level efforts.
- LANL identifies through equilibrium and bench scale testing a once-through 90% CO_2 capture promoter that supports the potential to achieve the Sequestration Program's cost target. Nexant is to perform an engineering analysis case to verify any economic benefits, as needed; no ETM validation is required, however, for this promoter for FY06.
- The CO_2 hydrate once-through process is to be validated at 1000 psia with the ETM at a CO_2 capture rate of 60% without H_2S. The performance of 68% rate of capture is based on a batch, equilibrium data with H_2S. Validation of the test results is required through multiple runs and engineering calculations. Operational issues will be solved that will specifically effect the validation of the technology.

Nexant was given the primary responsibility for Target #1, while Simteche was mainly responsible for Target #2; with LANL having the responsibility of Targets #3 and #4.

Six (6) engineering analysis tasks were identified to be performed by Nexant, with support from Simteche and Los Alamos National Laboratory (LANL) as required. These tasks were funded under a contract modification #017 to the Phase II work. The six tasks were:

1. Replace ammonia vapor compression refrigeration with ammonia absorption refrigeration to lower refrigeration costs,
2. CO_2 capture with promoted solvent,
3. Supercritical CO_2 recovery,
4. 3,000 psia Hydrate Formation Reactor Operating Pressure,
5. Recycle hydrate slurry to provide better control of reactor heat removal, and
6. Promoters that enable reactor heat removal using cooling water instead of ammonia refrigeration.

After completing and reviewing the results of the first four tasks, and with new laboratory data from LANL and Simteche, Nexant received approval of replacing the objective of Task #5 and #6 to investigate the following:

- Intermediate 1800 psia Hydrate Formation Reactor Operating Pressure (Task 5).
- 2-Step Hydrate Formation Reactor at elevated operating pressure (Task 6).

All six engineering analysis tasks were completed, with individual reports issued to DOE for review and comment. The current report presents a brief summary of the objective, results and conclusion for each of the six engineering analyses performed. Individual task reports, which contain more technical details, are attached in the Appendices.

Table 1 (page 3-7) summarizes the overall engineering economic analysis for five out of the six case studies. Case #1 (i.e, Task #1) results are not included as it was determined early that ammonia absorption has no cost advantages over ammonia vapor compression as the refrigeration process. These results are compared and tabulated side-by-side with the base case IGCC design without any CO_2 capture, a design with CO2 capture using Selexol and a 2-stage promoter hydrate process that were reported in the Phase II Engineering Analysis Report, issued in February 2006. Detailed case process design, heat and material balances, sized equipment and cost estimation can be found in the individual task reports. Economic assumptions and methodology used are the same as reported in the Phase II Engineering Analysis study.

As noted in the table, there were few minor adjustments made to the Selexol case results. The Phase II Engineering Analysis excluded some process steam requirement for the Selexol process, as well as small amount of power needed for the molecular sieve regeneration. These have been corrected, and as a result, the net power produced for this case was reduced from 391 to 384 MW. This reduction increased the COE from 6.13 to 6.24 cents/kWhr.

Table 1 compared the various IGCC case studies, in terms of percentage of CO_2 capture, net power production, COE (cost of electricity) and its increase with CO_2 capture, and the estimated CO_2 captured (avoided) cost, defined as $[COE_c - COE_r]/[CO2_r - CO2_c]$ where

COE_c is the COE for the IGCC plant with CO_2 capture, in $/MWhr
COE_r is the COE for the reference IGCC plant with no CO_2 capture, in $/MWhr
$CO2_r$ is the CO_2 emitted for the reference IGCC plant with no CO_2 capture, in tons/MWhr, and
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As shown in table 1, the best economics for a process that recovers 90% CO_2 in a single stage operation is between the conventional Selexol process and the Simteche 2-Step Hydrate Formation Reactor of Task #6. Selexol has a slightly lower COE of 6.24 cents/ kWhr versus 6.36 cents for the Simteche process. The 2-Step Hydrate Formation Reactor design of Task #6 represents a new concept of Simteche that would require experimental verification. The information is PROPRIETARY and patent pending.

To meet 90% CO_2 capture within a single-stage process, the 2-Stage Promoter Simteche process has the lowest COE and CO_2 capture costs, but this case was based on a theoretical hydrate structure with a low hydration number that has not yet been achieved (proven) at LANL. Low hydration number results in a lower heat of formation and thus lower capital and operating costs.

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Section 3 Brief Summary of Task Objective, Results and Conclusion

3.1 TASK 1 – REPLACING AMMONIA VAPOR COMPRESSION REFRIGERATION WITH AMMONIA ABSORPTION REFRIGERATION TO LOWER REFRIGERATION COSTS

Objective – This task has the objective of determining if ammonia absorption refrigeration can supply the refrigeration requirements for the Simteche CO_2 removal process more cost effectively than with ammonia vapor compression refrigeration. The refrigeration compressor is a major cost item with the ammonia vapor compression refrigeration process, which is not required with an absorption process.

Results and conclusion – Techno-economic analysis comparing ammonia absorption vs vapor compression refrigeration was performed on a fully integrated IGCC (EPRI prototype for the project) plant with the Simteche's non-promoted Once-Through CO_2 hydrate capture process. This was done because in addition to potential capital cost difference between the two ammonia refrigeration processes, there is significant difference in utility requirement, which would impact the overall IGCC plant CO_2 capture economics.

Engineering analysis shows that while the Ammonia Absorption does have an advantage in capital cost saving (about $7.7MM), it is offset by a higher utility requirement (i.e., low pressure steam that would need to be extracted from the IGCC plant) resulting in (a) lower net power production (by about 25.1 MW) and (b) higher cost of electricity production (6.18 instead of 5.75 cents per kWhr) making it less attractive as an alternative for hydrate refrigeration.

3.2 TASK 2 – CO2 CAPTURE WITH PROMOTED SOLVENT

Objective – This task investigated the effect of a promoter (e.g., tetrahydrofuran or THF) on the Simteche CO_2 capture process using a one-stage hydrate reactor. It has been observed at LANL (R. Currier et. al., 5th Annual Conference on Carbon Capture & Sequestration, May 2006) that a promoter can significantly alter the phase envelope of CO_2 hydrate formation, by increasing the HFT (maximum hydrate formation temperature) at which CO_2 hydrate formation can occur at a given CO_2 partial pressure to a higher temperature. This would allow the Simteche process to be operated at higher temperatures, which may result in potential savings in refrigeration cost. In a previous study (Phase II Engineering Analysis report of February 2006), the effect of promoter was investigated in a two-stage process.

The objectives of the current task were to (1) develop a single-stage promoter process design, completed with heat, material and utility balances, and capital cost estimates to allow its techno-economic feasibility to be assessed, and (2) compare the CO_2 capture cost of this process with the two-stage promoter process and with the Selexol process reported in the Phase II Engineering Analysis Report.

Results and conclusion – The engineering analysis equipped with process scheme, detailed heat and material balance was developed for this case based on the assumption that 90% CO_2 capture is achievable with the use of a promoter solvent. A ‘SII structure1’ type of CO_2 clathrate with a hydrate number of 8.5 and a heat of hydrate formation of 89.20 kJ/gmole CO_2 were assumed. These assumption would need to be verified once the promoter hydrate equilibrium correlations are developed based on the work from LANL.

Results indicate that while the addition of a promoter additive allows the Simteche CO_2 capture process to be operated at a higher temperature, the CO_2 hydrate formed has a different clathrate structure with a higher hydration number and heat of formation (without a promoter, the clathrate has a hydration number of 6.0 instead of 8.5 and a heat of formation of 60.0 instead of 89.20 KJ/gmole). And as a result, significantly higher refrigeration loading and ammonia refrigeration compression horsepower are required for the process, making it even less competitive when compared to the Selexol process. More details can be found in the individual task report of Appendix B.

3.3 TASK 3 – SUPERCRITICAL CO2 RECOVERY

Objective – This task has the objective of determining if pumping the hydrate slurry to a supercritical condition of 2200 psia is a cost effective way of eliminating the CO_2 sequestration compression costs. The CO_2 sequestration compressor is a major cost item and source of energy consumption for the Simteche CO_2 removal process. The process involves pumping the liquid CO_2 hydrate slurry produced from the Hydrate Formation Reactor to 2200 psia, and then recover a supercritical CO_2 stream from the Hydrate Flash Reactor, suitable for sequestration, without any additional compression.

Results and conclusion – Engineering analysis showed that there is no significant savings in pumping the CO_2 hydrate formed from the Hydrate Formation Reactor to its sequestration pressure of 2000 psi prior to CO_2 regeneration from the Hydrate Flask Reactor. The reasons are twofold: First, process simulation and engineering analysis showed that because of the proximity in density between the supercritical CO_2 and water, effective CO_2 recovery from the Hydrate Flask Reactor would have to be operated at close to 100 ^oF and then cooled the recycled water back to 34 ^oF. As a result, the overall refrigeration load of the system increases significantly. Secondly, it was found that that the energy required to pump the hydrate slurry to supercritical conditions was almost as much as that required for compressing the CO_2 by itself after it had been generated from the Hydrate Flask Reactor. This is because the hydrate slurry contains significantly more mass (~ 6 times more) than CO_2 by itself. Also, with CO_2 compression, once it is compressed to its supercritical condition of ~ 1100 psi, it can be pumped to 2200 psi as a ‘liquid’ via a CO_2 supercritical pump at relatively low power requirement. More details can be found in the individual task report of Appendix C.

3.4 TASK 4 – 3000 PSIA HYDRATE FORMATION REACTOR OPERATING PRESSURE

Objective – It is theoretically possible to achieve a 90% CO_2 capture for the Simteche process in a single stage by increasing the CO_2 partial pressure in the syngas feed. Preliminary calculations by Simteche have shown that this can be achieved at a hydrate reactor operating condition of 3,000 psia and 34 °F.

The objective of this task is to revise the process scheme to allow the hydrate formation reactor to be operated at 3000 psia and 34°F and to determine the capital and operating costs for the process. The results can then be compared to the 2-stage promoter case operation and also to the optimized Selexol process which are designed to recover 90% CO_2.

Results and conclusion – There is a considerable increase in both operating and capital costs associated with this process scheme of operating the hydrate formation reactor at 3000 psia. The increase in costs makes the scheme economically less attractive, when compared with previous cases. The cost increases are mostly due to the following:

- Hydrate formation reactor has to be operated at three times the pressure at which syngas is required to be delivered at the outlet of the plant,
- Higher operating pressure resulted in more expensive reactors, exchangers and vessels,
- Additional equipment such as two stages of high pressure compression and expansion are required,
- The recycle water pumping energy also became significant, and
- CO_2 must be recovered at the same low pressures as the base case, so there are no savings in CO_2 compression costs from the base case

When compared with the previous case studies of the 2-stage promoter scheme and the Selexol case, this 3000 psia scheme has a higher calculated cost of electricity (COE) - at 6.65 cents per kWhr vs. 6.13 and 6.24 cents/kWhr for the 2-stage promoter and Selexol case respectively. It represents a 45% increase in COE when compared to the base case IGCC with no CO_2 capture. More details can be found in the individual task report of Appendix D.

3.5 TASK 5 – 1800 PSIA HYDRATE FORMATION REACTOR OPERATING PRESSURE

Objective – The results of Task 4 (3000 psia hydrate formation reactor operating pressure case) indicates that considerably higher capital and operating costs will be incurred if the Simteche process is to capture 90% of the CO_2 in a single pass without the addition of a promoter. An optimum recovery for the Simteche process is probably less than 90%. A logical pressure to operate the Hydrate Formation reactor is probably around 1800 psia which can be achieved with a single stage syngas compressor from either a low pressure (500 psig) gasifier or from a high pressure (1050 psig) gasifier. Experimental observations at LANL show that hydrate formation can be greatly accelerated by increasing the CO_2 partial pressure and Simteche preliminary calculation also showed that an 84% CO_2 recovery can be achieved at hydrate reactor operating conditions of 1,800 psia and 34°F.

The objective of this task is revise the process scheme to allow the hydrate formation reactor to be operated at the intermediate pressure of 1800 psia and 34°F and to determine the capital and operating costs for the process. In addition, using the capital and operating costs from the 1800 psia and 3000 psia hydrate reactor operating conditions, a trend curve can be plotted showing the effect on cost of increasing CO_2 recovery with the Simteche process. This can then be compared to the base case operation at 1,000 psia and 34^{o}F and also to the optimized Selexol process which is designed to recover 90% CO_2.

Results and conclusion – The process design was revised, heat and material balances calculated and both capital and operating costs re-estimated. At 1800 psia hydrate formation reactor operating pressure, only 84% CO_2 capture can be achieved in a single stage reactor. As with the 3000 psia case of increasing the feed syngas pressure, substantial increases in both capital and operating costs resulted, due to:

- Higher cost for equipment associated with high pressure operation,
- Additional equipment requirement such as two stages of high pressure compression and expansion are required,
- Significant energy loads for the recycle water pumps, and
- No savings in CO_2 compression costs compared to the 1000 psia base case design as the CO_2 must be recovered at the same low pressure.

When compared with the previous case studies of the 2-stage promoter scheme and the Selexol case (both with 90% capture), this 1800 psia scheme has a somewhat lower cost of electricity (COE) - at 6.17 cents per kWhr vs. 6.13 and 6.24 cents/kWhr for the 2-stage promoter and Selexol case respectively. It represents a 35% increase in COE when compared to the base case IGCC with no CO_2 capture. More details can be found in the individual task report of Appendix E.

3.6 TASK 6 – TWO-STEP 2400 PSIA HYDRATE FORMATION REACTOR OPERATING PRESSURE

Objective – The results of Task 4 (3000 psia hydrate formation reactor operating pressure) indicates that considerably higher capital and operating costs will be incurred if the Simteche process is to recover 90% of the CO_2 in a single stage hydrate formation reactor. A way to reduce this cost is to utilize a scheme with a two-step hydrate formation reactor without any separation equipment between them but operating at different temperatures. In this way, a first step of bulk hydration formation can be achieved at a higher temperature followed by the final removal in a lower temperature reactor. This will reduce the refrigeration loads and as a result reduce the capital and operating costs. Experimental observations at LANL have shown that hydrates form at considerably higher temperatures when the reactors are operated at higher CO_2 partial pressures.

This 'Two-Step Hydrate Formation' is the latest concept developed by Simteche. Preliminary back-of-the-envelope calculations by Simteche shows that 68% of the Hydrates can be recovered at 7^{o}C and the other 22% can be captured at 0.5^{o}C (for a total of 90%) if the Hydrate reactor is operated at around 2400 psia. This higher operating pressure can be achieved by operating the gasifier at as high a pressure as possible, supplemented by additional syngas compression.

The objective of this task is to revise the Simteche CO_2 Hydrate process scheme allowing the first section of the hydrate reactor to be operated at 7^{o}C and 2400 psia, followed by a second reactor section at 0.5 ^{o}C and 2360 psia; then carry out design heat, material and utility balance calculations, and cost estimation to allow the technical and economic analyses of this new process scheme.

Results and conclusion – The capital cost and energy consumption of this 'Two-Step Hydrate Formation' scheme is the lowest of all the Simteche schemes studied thus far that can remove 90% CO_2 in a single stage. When compared with the previous case studies of the 2-stage promoter scheme and the Selexol case (both also with 90% capture), this 'Two-Step Hydrate Formation' scheme has a cost of electricity (COE) - at 6.36 cents per kWhr vs. 6.13 and 6.24 cents/kWhr for the 2-stage promoter and Selexol case respectively. It represents a 39% increase in COE when compared to the base case IGCC with no CO_2 capture. More details can be found in the individual task report of Appendix F.

3.7 OVERALL ENGINEERING ANALYSIS SUMMARY

Table 1 summarizes the overall engineering economic analysis of the five out of the six case studies. Case #1 (i.e, Task #1) results are not included as it was determined early that ammonia absorption has no cost advantages over ammonia vapor compression as a refrigeration process. These results are compared and tabulated side-by-side with the base case IGCC design without any CO2 capture, a design with CO2 capture using Selexol and a 2-stage promoter hydrate process that were reported in the Phase II Engineering Analysis Report, issued in February 2006. Detailed case process design, heat and material balances, sized equipment and cost estimation can be found in the individual task reports. Economic assumptions and methodology used are the same as in the Phase II Engineering Analysis study.

As noted in the table, there were some minor adjustments made to the Selexol case results. The Phase II Engineering Analysis excluded some process steam requirement for the Selexol process, as well as small amount of power needed for molecular sieve regeneration. These have been corrected, and as a result, the net power produced for this case is reduced from 391 to 384 MW. This reduction increased the COE from 6.13 to 6.24 cents/kWhr.

Table 1 compared the various IGCC case studies, in terms of percentage of CO_2 capture, net power production, COE (cost of electricity) and its increase with CO_2 capture, and the estimated CO_2 captured (avoided) cost, defined as $[COE_c - COE_r]/[CO2_r - CO2_c]$ where

COE_c is the COE for the IGCC plant with CO_2 capture, in $/MWhr
COE_r is the COE for the reference IGCC plant with no CO_2 capture, in $/MWhr
$CO2_r$ is the CO_2 emitted for the reference IGCC plant with no CO_2 capture, in tons/MWhr, and
$CO2_c$ is the CO_2 emitted for the IGCC plant with CO_2 capture, in tons/MWhr.

As shown in table 1, the best economics for a process that recovers 90% CO_2 in a single stage operation is between the conventional Selexol process and the Simteche 2-Step Hydrate Formation Reactor of Task #6. Selexol has a slightly lower COE of 6.24 cents/ kWhr versus 6.36 cents for the Simteche process. The 2-Step Hydrate Formation Reactor design of Task #6 represents a new concept of Simteche that would require experimental verification. The information is PROPRIETARY and patent pending.

To meet 90% CO_2 capture within a single-stage process, the 2-Stage Promoter Simteche process has the lowest COE and CO_2 capture costs, but this case was based on a theoretical hydrate structure with a low hydration number that has not yet been achieved (proven) at LANL. Low hydration number results in a lower heat of formation and thus lower capital and operating costs.

Table 1 – Engineering Analysis Summary

Task	2005 (1) Base Case IGCC	2005 (1), (2) IGCC with Selexol	SIMTECHE PROCESS 2005 (1) Promoter 2-Stage Promoter	2006 Task 2 1-Stage Promoter	2006 Task 3 Supercritical CO2	2006 Task 4 1-Step 3000 Psia Reactor	2006 Task 5 1-Step 1800 Psia Reactor	2006 Task 6 Adv. 2400 Psia Reactor
CO_2 Separation Ratio	**0%**	**90%**	**90%**	**90%**	**68%**	**90%**	**84%**	**90%**
Onstream Factor	80%	80%	80%	80%	80%	80%	80%	80%
Coal Feed (as received), Short Tons per Year	1,009,152	1,118,068	1,118,068	1,118,068	1,118,068	1,118,068	1,118,068	1,118,068
Carbon Dioxide Captured, STPY		2,288,719	2,278,145	2,278,499	1,724,241	2,297,500	2,120,403	2,273,528
Carbon Dioxide Emitted, STPY	2,353,363	237,382	247,995	252,624	801,909	228,640	2,118,272	252,612
Cost of Production, $ per Year								
Capital Charge	$ 87,113,705	$ 112,484,409	$ 115,290,529	$ 119,706,087	$ 115,287,317	$ 119,581,526	$ 115,651,285	$ 116,829,320
Coal Cost	$ 29,265,408	$ 32,423,972	$ 32,423,972	$ 32,423,972	$ 32,423,972	$ 32,423,972	$ 32,423,972	$ 32,423,972
Chemicals/Consumables	$ 2,271,894	$ 2,353,786	$ 2,285,469	$ 2,285,469	$ 2,485,905	$ 2,282,874	$ 2,282,874	$ 2,282,874
Maintenance Cost, Materials&Labor	$ 9,218,382	$ 11,903,112	$ 12,200,056	$ 12,667,311	$ 12,199,716	$ 12,654,130	$ 12,238,231	$ 12,362,891
Operating Labor and Administration	$ 8,688,493	$ 8,830,182	$ 8,830,182	$ 8,830,182	$ 8,830,182	$ 8,830,182	$ 8,830,182	$ 8,830,182
Total Cost of Production, $ per year	**$ 136,557,882**	**$ 167,995,461**	**$ 171,030,207**	**$ 175,913,021**	**$ 171,227,092**	**$ 175,772,684**	**$ 171,426,545**	**$ 172,729,239**
Net Power Produced, kW_e	424,540	384,081	398,249	382,586	401,575	377,027	396,208	387,544
Cost of Electricity, cents per kWhr	4.59	6.24	6.13	6.56	6.08	6.65	6.17	6.36
% Increase in COE		36%	34%	43%	33%	45%	35%	39%
Avoided Capture Cost, $/Ton of CO_2	**$ -**	**$ 25.3**	**$ 23.6**	**$ 30.6**	**$ 32.8**	**$ 31.6**	**$ 26.7**	**$ 27.4**
Avoided Capture Cost, $/Ton of Carbon	**$ -**	**$ 6.9**	**$ 6.4**	**$ 8.3**	**$ 8.9**	**$ 8.6**	**$ 7.3**	**$ 7.5**

Note (1) - Case Studies from Phase II Engineering Analysis Report of February 2006
Note (2) The Selexol case has been adjusted for steam used in the reboiler; Phase II Engineering Analysis Report cited numbers excluding the steam usage.
Task 1 involved a refrigeration cycle trade-off analysis of using ammonia absorption which determined not to be economical.

Appendix A Task #1 - Comparison of Ammonia Vapor Compression Refrigeration with Ammonia Absorption Refrigeration

Task 1 - Comparison of ammonia vapor compression refrigeration with ammonia absorption refrigeration

Objective

This task has the objective of determining if ammonia absorption refrigeration can supply the refrigeration requirements for the Simteche CO_2 removal process more cost effectively than with ammonia vapor compression refrigeration. The refrigeration compressor is a major cost item with the ammonia vapor compression refrigeration process, which is not required with an absorption process. Absorption refrigeration has been used for more than 100 years. Most of the applications use lithium bromide. However Lithium Bromide is limited to refrigeration temperatures above 32 ^oF. For the Simteche CO_2 removal process, 20^oF is required. Ammonia absorption refrigeration can achieve as low as -40 ^oF in temperature.

Summary

Techno-economic analysis comparing ammonia absorption vs vapor compression refrigeration was performed on a fully integrated IGCC (EPRI prototype for the project) plant with the Once-Through Simteche's CO_2 hydrate capture process. This was done because in addition to potential capital cost difference between the two ammonia refrigeration processes, there is significant difference in utility requirement, which would impact the overall IGCC plant CO_2 capture economics.

Table 1-1 summarizes the major differences, and the final economic comparison on a cost of electricity basis. As can be seen, although the Ammonia Absorption Case has lower installed capital cost (about $7.7MM), it has a higher cost of electricity of 6.18 cents per kWhr compared to the 5.75 cents/kWhr than the compression refrigeration case. The process requires a significant level of low-pressure steam, which in the EPRI prototype IGCC design, was used to generate 43 MW of power. Tables 1-2 to 1.5 show that the ammonia absorption case has higher avoided capture cost of CO2 ($32 to 25/ton CO2) and lower overall energy efficiency (35.4 to 37.7%) than the ammonia compression refrigeration case. Because of these factors, it is concluded that there will be no achievable savings by using ammonia refrigeration process for the Simteche process. There may be other situations where excess low level waste heat is available where the conclusion could be different but not with EPRI prototype IGCC design.

Ammonia absorption refrigeration

The process utilizes water as a working fluid to absorb ammonia vapor from its evaporative refrigeration stage. Ammonia vapor is absorbed by weak aqueous ammonia solution in a water cooled absorber resulting in a strong liquid aqueous ammonia stream. The strong aqueous ammonia liquid is then pumped up to a high pressure distillation tower where heat is added and a high purity (99.96%) ammonia liquid is recovered from the tower overhead. Reboiler heat can be supplied using a low-level waste heat. A weak aqueous ammonia stream from the desorber tower bottoms is cooled and recycled to the ammonia absorber.

The primary advantage of the ammonia absorption refrigeration when compared against the vapor compression refrigeration system is in the ability to substitute the more costly ammonia compressors with the less expensive heated regenerator/ CW (circulating water) cooled absorber system. Also, the NH_3 absorption system has fewer moving parts, thus requiring less maintenance than the mechanical compressor driven vapor re-compression system. These advantages depend on the availability of low-level waste heat for the regeneration of high purity ammonia from the aqua ammonia mixture. For the level of refrigeration required by the Simteche process (in the 20 ^{o}F range), the heat source required for regeneration is around 250 to 260 ^{o}F, which is equivalent to 30 to 35 psia steam.

The lowest usable steam available in the EPRI IGCC prototype plant is the 58 psia, 600 ^{o}F LP steam. This is currently used for power recovery exhausting at 2" through the LP case of the condensing steam turbine. To maximize the low pressure steam value, the 58 psia 600 ^{o}F steam can be sent to an extraction turbine and exhaust at 35 psia (the required pressure level for the desorber reboiling) and generate some additional power.

Methodology

There are several major suppliers that provide large size ammonia absorption units in the 150 to 5000 tons of refrigeration range. These include:

- Toromont Energy Systems-Canada and US
- Transparent Energy Systems- India
- Colibri BV – Netherlands

Toromont was contacted, but indicated they hadn't built a unit in 20 years and are no longer pursuing this business. Their current market focus is on Lithium bromide absorption refrigeration which is limited to refrigeration temperature above 32^{o}F. The other vendors were contacted, but they have not responded to requests for capital and operating costs.

As a result of lack of vendor input, a simulation model for a single stage ammonia absorption plant was built using HYSYS simulator, to size and cost the process equipment, along with estimating the overall process utility (e.g., temperature level of regeneration steam required) consumption. Figure 1 shows the PFD for the ammonia absorption process. The model was first calibrated against literature published heating and cooling requirements (including values in the vendor brochures), before being used for the current study. With the model, the desorber pressure was set by the condensing temperature of ammonia at the top of tower. The bottom temperature was determined by the concentration of the weak ammonia solution used in the water cooled absorber. The strong aqueous ammonia solution concentration from the absorber was set by the cooling water temperature approach in the water cooled absorber. The ammonia pickup and the difference in ammonia concentrations of the aqueous solutions determine the aqueous ammonia circulation rate.

In addition, the desorber overhead temperature was set at 95 ^{o}F in order to give a reasonable approach to the cooling water used to condense the overhead. The desorber bottom temperature of 240 ^{o}F resulted in the need to use a 35 psia steam in order to provide reasonable size reboilers for the desorber.

Same refrigeration (i.e., hydration formation reactors, hydrate flash reactor A & B, syngas chilling and recycled water chilling) duties were used as in the February 2006 Engineering Analysis report case of 'Simteche Once-Through' Design.

Four (4) separate ammonia absorption trains are required.

A similar simulation was performed for the ammonia compression refrigeration design, to ensure that the comparison was done on a consistent basis. The results are shown in the following:

Table 1-1 - Economic Comparative Analysis
Table 1-2 - Equipment Cost Estimates
Table 1-3 - Power Production Summary
Table 1-4 - Carbon Control Cost
Table 1-5 – IGCC Plant Capital Cost
Table 1-6A - Equipment list Ammonia Absorption Refrigeration
Table 1-6B - Equipment list Ammonia Compression Refrigeration
Figure 1 - Simplified PFD Ammonia Absorption Refrigeration
Figure 2 - Simplified PFD Ammonia Compression Refrigeration
Table 2.1 – Utility summary (Ammonia absorption)
Table 2.2 – Utility summary (Ammonia compression)

Results

In sizing the equipment for the Ammonia Absorption case, it became apparent that multiple trains are required due to the physical size of the equipment. In order to accommodate the desorber reboilers at the base of the desorber, it was necessary to have four (4) processing trains, resulting in an overall capital cost of $54.4 million.

The corresponding capital cost for the Ammonia Compression design is $58.2 million of which, as expected, the centrifugal compressor is a major cost item.

The utility requirements for the two refrigeration case scenarios are very different, which significantly impacted their comparative economics. A key difference is that the Ammonia Absorption Case requires a significant quantity of low pressure steam, and despite the saving in auxiliary power loading of not having to run a compressor, the overall integrated IGCC plant generates less power (~ 26MW), resulting in a higher cost of electricity. And, because of the large low pressure steam requirement, the integrated IGCC plant requires an additional small low-pressure steam boiler, along with the associated make-up fuel requirement and higher cooling water loading.

The Ammonia Compression case requires a surface condenser on the low pressure steam turbine, which is not needed for the Ammonia Absorption case; the Ammonia Compression low pressure turbine is larger than that in the Ammonia Absorption case since it produces 43 MW more power. Since there was less low temperature condensate in the Ammonia Absorption case, additional cooling was required in the gasifier final cooling section.

Tables 1-1 to 1-5 summarize the major differences, and the final economic comparison on a cost of electricity basis. As can be seen in Tables 1-1 &1-5, while the ammonia absorption case has a slight advantage in installed capital cost saving of $7.66 million, it is economically less attractive because of the overall loss in power production from the low-pressure steam turbine within the IGCC design. It results in its cost of electricity being 6.18 compared to 5.75 cents per kWhr for the ammonia compression case. Tables 1.3 and 1.4 show that the ammonia absorption case has higher avoided capture cost of CO_2 ($32 to 25/ton CO_2) and lower overall energy efficiency (35.4 to 37.7%) than the ammonia compression refrigeration case. However, ammonia absorption has a lower parasitic electrical load (2.6 to 6.0%) than ammonia compression since it has a lower electrical requirement by eliminating the compression electrical load.

With four processing trains, the Ammonia Absorption plant is expected to be more complex and would take additional staff to operate and maintain. As cited earlier, because insufficient low-pressure steam is available from the design IGCC plant for absorption regeneration; an additional LP steam boiler is required. Alternatively, one can add duct firing to the IGCC HRSG. Under a different scenario (i.e., if a source of waste heat was available that wasn't already earmarked for power generation), the study conclusion might very well be different.

Table 1-1
Economic Comparison Analysis

Case		Once-Thru Case with NH3 Compression Refrigeration	Once-thru Case with NH3 Absorption Refrigeration
CO2 Recovery		68%	68%
Total Installed IGCC Plant Cost	$MM	736.9	729.3
Power Revenues			
Power Generation			
Gas Turbine Power	KW	342,295	342,295
HP & IP Steam Turbine	KW	98,263	98,263
LP Steam Turbine	KW	52,051	9,094
Generator Loss	KW	(7,265)	(7,265)
Turboset Power	KW	485,344	442,387
Expander	KW	10,835	10,835
Total Gross Power 'Generated	KW	496,179	453,222
Auxiliary Loads	KW		
NH3 Refrigeration Compressors	KW	19,958	
NH3 Absorption Refrig	KW		1,982
Cooling Water Pumps	KW	3,879	4,421
Cooling Tower Fans	KW	1,141	1,296
Balance of Plant	KW	59,491	58,949
Total Auxiliary loads	KW	84,469	66,648
OnStream Factor		0.8	0.8
Net Power	KW	411,710	386,574
Cost of Electricity	Cents/kWhr	5.75	6.18

Table 1-2 Equipment Cost Estimates

CO_2 Removal and Compression	SIMTECHE	
	Once-Thru NH_3 Comp	***Once-Thru NH_3 Absorption***
CO_2 Separation Ratio	68%	68%
Total Installed Cost, $ 2Q 2005		
Vessels, Exchangers, Pumps	$ 32,227,217	$ 54,772,286
Refrigeration Compression	$ 12,999,977	$ -
CO_2 + Other Compression	$ 5,795,199	$ 860,567
MDEA AGR	$ 584,000	$ 584,000
Column Packing, Initial Fill	$ 24,521	$ 24,521
Total CO2 Removal and Compression	**$51,630,914**	**$56,241,374**

Table 1-3 Power Production Summary

Power Production Summary	**No Capture**	**SIMTECHE**	
		Once-Thru NH_3 Comp	***Once-Thru NH_3 Absorption***
CO_2 Separation Ratio	0%	68%	68%
Gross Plant Power, kW_e	474,040	496,179	453,222
Auxilliary Power Loads			
CO_2 Capture and Compression		29,648	11,672
Power Plant	49,500	54,821	54,976
Net Plant Power, kW_e	424,540	411,710	386,574
Net Efficiency, %HHV	43.0	37.7	35.4
Heat Rate, BTU/kWhr	7,936	9,041	9,629
Capture Parasitic Load, %		6.0%	2.6%

Table 1-4 Carbon Control Cost

Carbon Control Costs	No Capture	SIMTECHE	
		Once-Thru NH_3 Comp	*Once-Thru NH_3 Absorption*
CO_2 Separation Ratio	0%	68%	68%
Onstream Factor	80%	80%	80%
Coal Feed (as received), Short Tons per Year	1,009,152	1,118,068	1,118,068
Carbon Dioxide Captured, STPY		1,724,192	1,724,192
Carbon Dioxide Emitted, STPY	2,353,363	801,947	801,947
Cost of Production, $ per Year			
Capital Charge	$ 87,113,705	$ 110,539,513	$ 109,390,698
Coal Cost	$ 29,265,408	$ 32,423,972	$ 32,423,972
Chemicals/Consumables/Fuel Gas	$ 2,271,894	$ 2,276,798	$ 5,306,798
Maintenance Cost, Materials&Labor	$ 9,218,382	$ 11,697,303	$ 11,575,735
Operating Labor and Administration	$ 8,688,493	$ 8,830,182	$ 8,830,182
Total Cost of Production, $ per year	**$136,557,882**	**$ 165,767,768**	**$ 167,527,386**
Net Power Produced, kW_e	424,540	411,710	386,574
Cost of Electricity, cents per kWhr	4.59	5.75	6.18
		25%	35%
Avoided Capture Cost, $ per Ton of CO_2	**$ -**	**$ 25**	**$ 36**
Avoided Capture Cost, $ per Ton of Carbon	**$ -**	**$ 6.8**	**$ 9.8**
Capture Cost of Shift	**$ -**	$ 7.80	$ 8.69

Table 1-5 IGCC Plant Capital Cost

Capital Cost Summary	**No Capture**	**SIMTECHE**	
		Once-Thru NH_3 Comp	***Once-Thru NH_3 Absorption***
CO_2 Separation Ratio	0%	68%	68%
Bare Erected Cost, $ 2Q 2005			
Gasifier/ASU	$178,727,027	$ 194,122,096	$ 194,122,096
Gas Cleanup, Shift and Saturation	$ 36,621,825	$ 83,534,325	$ 83,534,325
CO_2 Removal and Compression		$ 51,630,914	$ 56,241,374
Gas Turbine and Steam Power Generation	$128,613,519	$ 128,572,504	$ 116,192,393
Balance of Plant	$116,956,704	$ 127,005,310	$ 128,696,576
Bare Erected Cost	$460,919,076	$ 584,865,149	$ 578,786,765
Home Office Cost	$27,655,145	$35,091,909	$34,727,206
Contingency	$92,183,815	$116,973,030	$115,757,353
Total Installed Cost (TIC)	**$580,758,036**	**$736,930,088**	**$729,271,323**

Table 1-6A Equipment List-Ammonia Absorption Refrigeration

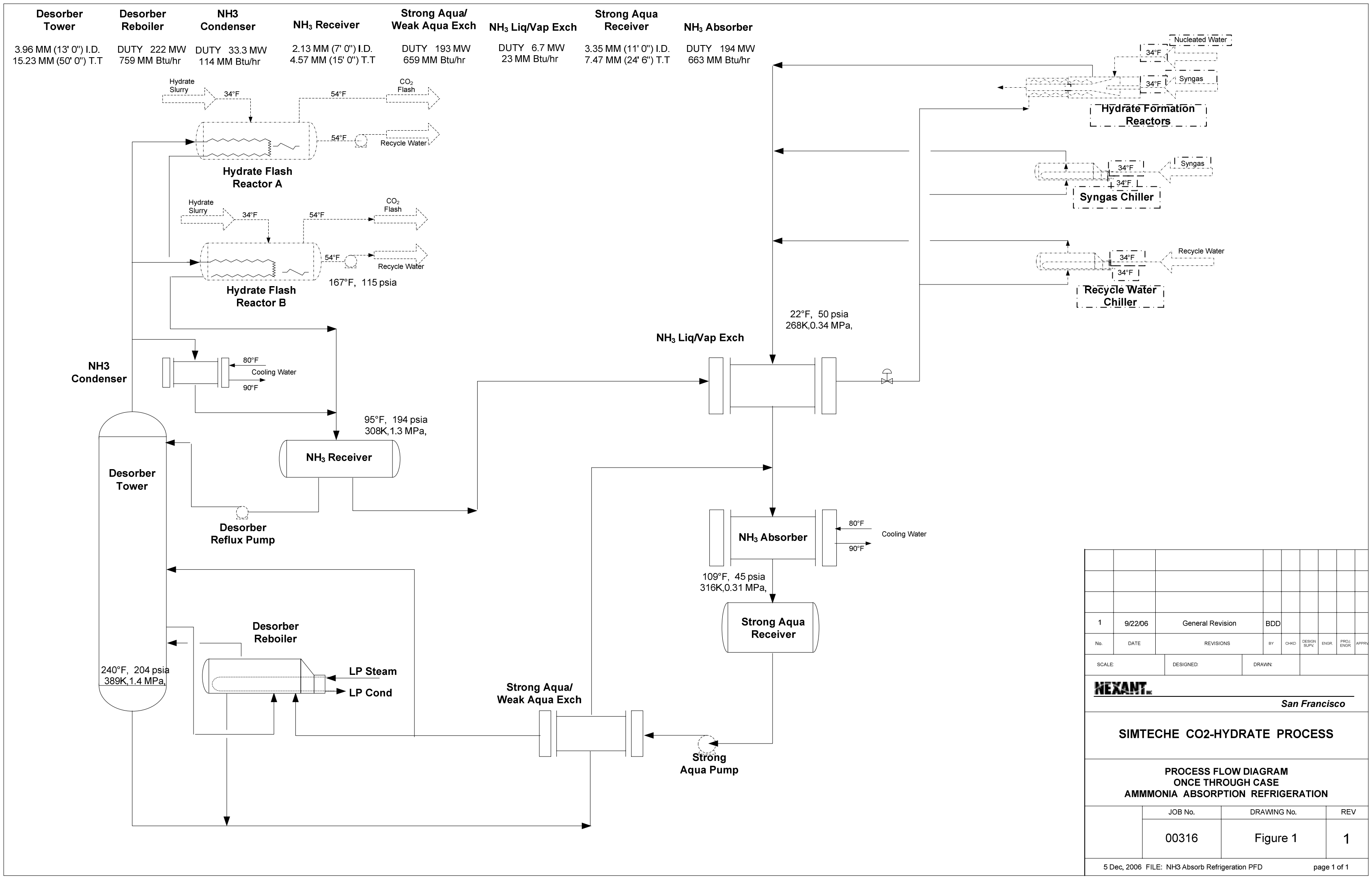
Equipment	No. ea.	Size/Duty	Material Tube/Shell	Temperature (°F) Tube/Shell	Pressure (psig) Tube/Shell
Reactors					
Hydrate Formation Reactors	10	329 MM Btu/hr total	304 SS/CS	200/200	1100/235
Hydrate Flash Reactor A	6	363.4 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/680
Hydrate Flash Reactor B	1	32.0 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/315
Exchangers					
Syngas Chiller	1	21.7 MM Btu/hr	304 SS/CS	200/200	1,100/235
Recycle Water Chiller	2	162. MM Btu/hr	304 SS/CS	200/200	300/235
Flash Gas Cooler	1	3.2 MM Btu/hr	CS/CS	285/200	680/50
CO_2 Sequestration Cooler	1	52.5 MM Btu/hr	CS/CS	375/200	2,400/50
Ammonia Condenser	8	113.7MM Btu/hr	CS/CS	200/360	500/235
Acid Gas Comp 1st Stg Intercooler	1	0.1 MM Btu/hr	CS/304 SS (Clad)	200/350	50/50
Acid Gas Comp 2nd Stg Intercooler	1	0.1 MM Btu/hr	CS/304 SS (Clad)	200/350	50/110
Acid Gas Comp 3rd Stg Intercooler	1	0.1 MM Btu/hr	CS/304 SS (Clad)	200/350	50/320
Regeneration Heater-Electric	1	200 kW	CS/CS	400	15
Liq NH3/Vap NH3 Exchanger	4	23.8M Btu/hr	CS/CS	300/300	215/50
NH3 Absorption Coolers	8	663.3M Btu/hr	CS/CS	300/300	100/50
Strong Aqua/Weak Aqua Exchanger	4	658.6M Btu/hr	CS/CS	300/300	215/225
Desorber Reboiler	16	758.5M Btu/hr	CS/CS	500/400	300/225
Gasifier Final Cooler	1	16.3M Btu/hr	CS/304 SS (Clad)	300/350	100/500
Vessels					
Syngas KO Drum	1	8'ID x 12' T-T	304 SS Clad	155	1,055
Slurry/Gas Separator	1	14ID x 40' T-T	304 SS Clad	100	985
Flash Gas Compr KO Drum	1	3.5'ID x 7' T-T	304 SS Clad	120	300
Acid Gas Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	50
Acid Gas 1st Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	100
Acid Gas 2nd Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	230
Acid Gas Product KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	670
Mol Sieve Dryer	2	6'ID x 12' T-T	304 SS Clad	450	670
Ammonia Receiver	4	7'ID x 15' T-T	CS	200	235
Strong Aqua Receiver	1	11'ID x 24.5' T-T	CS	200	235
Desorber Tower	4	13'ID x 24.5' T-T with/8 valve trays	CS with 410 SS valve trays	350	225
Compressors					
Flash Gas Compressor	1	1,415 HP	CS	350	625
CO2 Sequestration Compressor	1	10,800 HP	CS	350	2400
Acid Gas Compressor	1	125 HP	CS	350	610
Pumps					
Chilled Water Recycle Pumps (3-50% pumps)	3	6,400 GPM/ 4,000 HP each	304 SS	200	1100
Cooling Water Pumps (3-50% pumps)	8	12,000 GPM/400 HP each	CS	200	100
Demineralized Water Pumps (2-100% pumps)	2	0.14 GPM/0.03 HP each	CS		

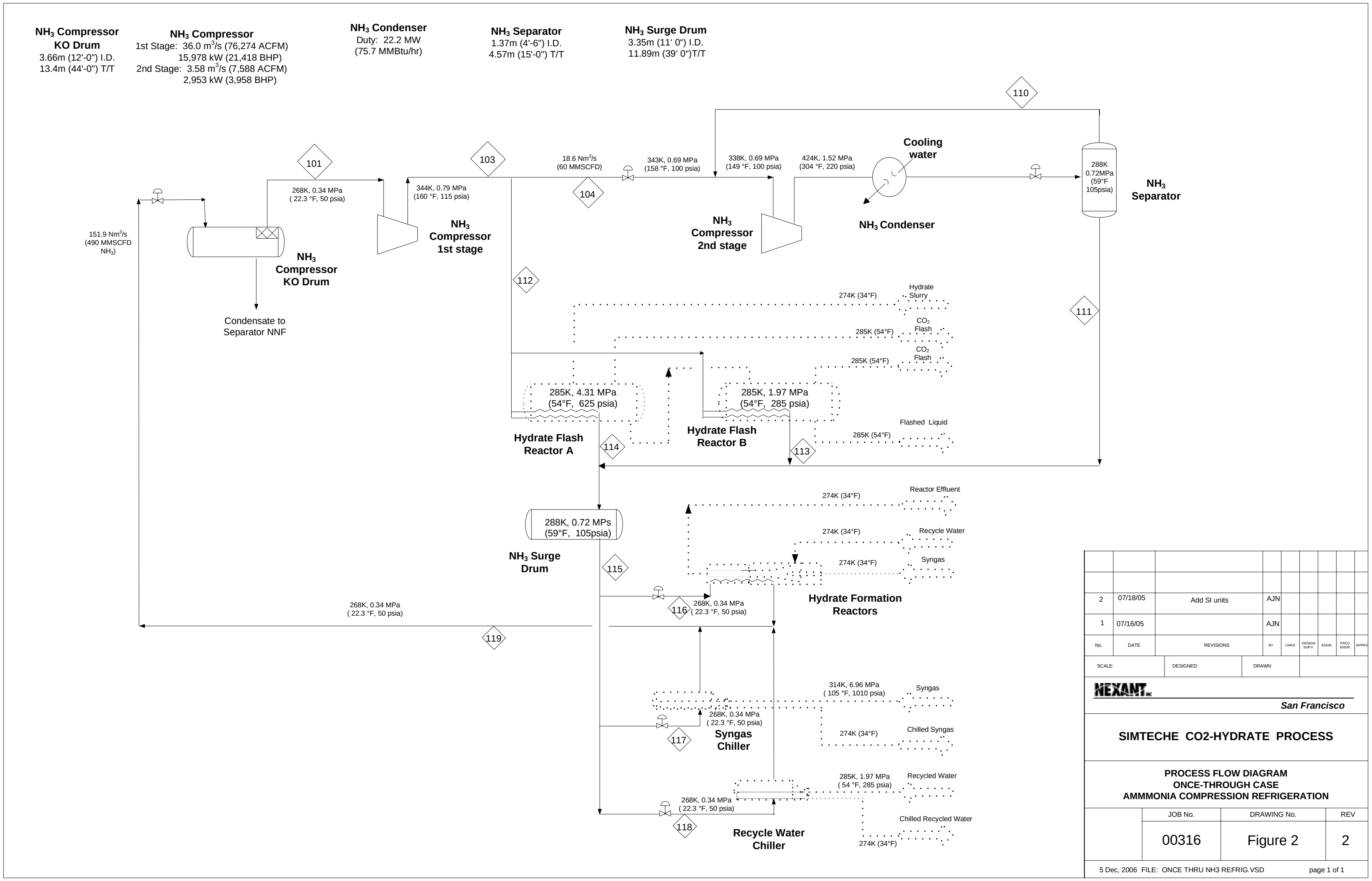
Equipment	No. ea.	Size/Duty	Material	Temperature (°F)	Pressure (psig)
			Tube/Shell	Tube/Shell	Tube/Shell
Lube Oil Pumps (3-50% pumps)	3	456 GPM/1 HP each	CS		
Strong Aqua Pump		4030 GPM/ 700 HP each	CS	200	100
NH3 Absorber Reflux Pump		4030 GPM/ 700 HP each	CS	200	100
Miscellaneous					
LP Condensing Steam Turbine		9100 KW 58 psia 600F to 35psia			
Cooling Tower		200,000 gpm, 1216 MM Btu/hr duty			
20 psig Steam Boiler		57,000 lb/hr 20 psig saturated steam			

Table 1-6B   Equipment List-Ammonia Compression Refrigeration

Equipment	No. ea.	Size/Duty	Material Tube/Shell	Temperature (°F) Tube/Shell	Pressure (psig) Tube/Shell
Reactors					
Hydrate Formation Reactors	10	329 MM Btu/hr total	304 SS/CS	200/200	1100/235
Hydrate Flash Reactor A	6	363.4 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/680
Hydrate Flash Reactor B	1	32.0 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/315
Exchangers					
Syngas Chiller	1	21.7 MM Btu/hr	304 SS/CS	200/200	1,100/235
Recycle Water Chiller	2	162. MM Btu/hr	304 SS/CS	200/200	300/235
Flash Gas Cooler	1	3.2 MM Btu/hr	CS/CS	285/200	680/50
CO_2 Sequestration Cooler	1	52.5 MM Btu/hr	CS/CS	375/200	2,400/50
Ammonia Condenser	2	179.3 MM Btu/hr	CS/CS	200/360	500/235
Acid Gas Comp 1st Stg intercooler	1	0.1 MM Btu/hr	CS/304 SS (Clad)	200/350	50/50
Acid Gas Comp 2nd Stg Intercooler	1	0.1 MM Btu/hr	CS/304 SS (Clad)	200/350	50/110
Acid Gas Comp 3rd StageIntercooler	1	0.1 MM Btu/hr	CS/304 SS (Clad)	200/350	50/320
Regeneration Heater-Electric	1	200 kW	CS/CS	400	15
Vessels					
Syngas KO Drum	1	8'ID x 12' T-T	304 SS Clad	155	1,055
Slurry/Gas Separator	1	14ID x 40' T-T	304 SS Clad	100	985
Flash Gas Compr KO Drum	1	3.5'ID x 7' T-T	304 SS Clad	120	300
Acid Gas Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	50
Acid Gas 1st Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	100
Acid Gas 2nd Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	230
Acid Gas Product KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	670
Mol Sieve Dryer	2	6'ID x 12' T-T	304 SS Clad	450	670
Ammonia Separator	1	6'ID x 17' T-T	CS	200	235
Ammonia Surge Drum	1	9.5'ID x 31' T-T	CS	200	235
Ammonia Compr KO Drum	2	13'ID x 39' T-T	CS	200	235
Compressors					
Flash Gas Compressor	1	1,415 HP	CS	350	625
CO2 Sequestration Compressor	1	10,800 HP	CS	350	2400
Acid Gas Compressor	1	125 HP	CS	350	610
NH3 Refrig 1st Stage Compressor	1	53,500 HP	CS	200	136
NH3 Refrig 2nd Stage Compressor	1	7,650 HP	CS	350	245
Pumps					
Chilled Water Recycle Pumps (3-50% pumps)	3	6,400 GPM/ 4,000 HP each	304 SS	200	1100
Cooling Water Pumps (3-50% pumps)	3	15,000 GPM/400 HP each	CS	200	100
Demineralized Water Pumps (2-100% pumps)	2	0.14 GPM/0.03 HP each	CS		
Lube Oil Pumps (3-50% pumps)	3	456 GPM/1 HP each	CS		

Equipment	No. ea.	Size/Duty	Material	Temperature (°F)	Pressure (psig)
			Tube/Shell	Tube/Shell	Tube/Shell
Miscellaneous					
LP Condensing Steam Turbine	1	52050 KW 58 psia to 2” HgA			
Cooling Tower	1	175600 GPM, 1108 MMBtu/hr duty			


Desorber Tower
3.96 MM (13' 0") I.D.
15.23 MM (50' 0") T.T
Desorber Reboiler
DUTY 222 MW
759 MM Btu/hr
NH3 Condenser
DUTY 33.3 MW
114 MM Btu/hr
NH3 Receiver
2.13 MM (7' 0") I.D.
4.57 MM (15' 0") T.T
Strong Aqua/ Weak Aqua Exch
DUTY 193 MW
659 MM Btu/hr
NH3 Liq/Vap Exch
DUTY 6.7 MW
23 MM Btu/hr
Strong Aqua Receiver
3.35 MM (11' 0") I.D.
7.47 MM (24' 6") T.T
NH3 Absorber
DUTY 194 MW
663 MM Btu/hr
Hydrate Slurry
34°F
54°F
CO2 Flash
Recycle Water
Hydrate Flash Reactor A
Hydrate Flash Reactor B
167°F, 115 psia
Nucleated Water
Syngas
Hydrate Formation Reactors
Syngas Chiller
Recycle Water Chiller
22°F, 50 psia
268K,0.34 MPa,
NH3 Liq/Vap Exch
NH3 Condenser
80°F
Cooling Water
90°F
95°F, 194 psia
308K,1.3 MPa,
NH3 Receiver
Desorber Tower
Desorber Reflux Pump
NH3 Absorber
109°F, 45 psia
316K,0.31 MPa,
Strong Aqua Receiver
Desorber Reboiler
240°F, 204 psia
389K,1.4 MPa,
LP Steam
LP Cond
Strong Aqua/ Weak Aqua Exch
Strong Aqua Pump
1
9/22/06
General Revision
BDD
No.
DATE
REVISIONS
BY
CHKD
DESIGN SUPV.
ENGR
PROJ. ENGR
APPR
SCALE:
DESIGNED:
DRAWN:
NEXANT
San Francisco
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
ONCE THROUGH CASE
AMMMONIA ABSORPTION REFRIGERATION
JOB No.
DRAWING No.
REV
00316
Figure 1
1
5 Dec, 2006 FILE: NH3 Absorb Refrigeration PFD
page 1 of 1



NH3 Compressor KO Drum
3.66m (12'-0") I.D.
13.4m (44'-0") T/T
NH3 Compressor
1st Stage: 36.0 m3/s (76,274 ACFM)
15,978 kW (21,418 BHP)
2nd Stage: 3.58 m3/s (7,588 ACFM)
2,953 kW (3,958 BHP)
NH3 Condenser
Duty: 22.2 MW
(75.7 MMBtu/hr)
NH3 Separator
1.37m (4'-6") I.D.
4.57m (15'-0") T/T
NH3 Surge Drum
3.35m (11' 0") I.D.
11.89m (39' 0")T/T
110
101
103
18.6 Nm3/s
(60 MMSCFD)
343K, 0.69 MPa
(158 °F, 100 psia)
338K, 0.69 MPa
(149 °F, 100 psia)
424K, 1.52 MPa
(304 °F, 220 psia)
Cooling water
288K
0.72MPa
(59°F
105psia)
NH3 Separator
268K, 0.34 MPa
(22.3 °F, 50 psia)
344K, 0.79 MPa
(160 °F, 115 psia)
104
151.9 Nm3/s
(490 MMSCFD
NH3)
NH3 Compressor 1st stage
NH3 Compressor 2nd stage
NH3 Condenser
NH3 Compressor KO Drum
Condensate to
Separator NNF
112
274K (34°F)
Hydrate Slurry
285K (54°F)
CO2 Flash
285K (54°F)
CO2 Flash
111
285K, 4.31 MPa
(54°F, 625 psia)
285K, 1.97 MPa
(54°F, 285 psia)
Flashed Liquid
285K (54°F)
Hydrate Flash
Reactor A
114
Hydrate Flash
Reactor B
113
274K (34°F)
Reactor Effluent
288K, 0.72 MPs
(59°F, 105psia)
274K (34°F)
Recycle Water
NH3 Surge Drum
274K (34°F)
Syngas
115
268K, 0.34 MPa
(22.3 °F, 50 psia)
116
268K, 0.34 MPa
(22.3 °F, 50 psia)
Hydrate Formation
Reactors
119
314K, 6.96 MPa
(105 °F, 1010 psia)
Syngas
268K, 0.34 MPa
(22.3 °F, 50 psia)
274K (34°F)
Chilled Syngas
117
Syngas
Chiller
285K, 1.97 MPa
(54 °F, 285 psia)
Recycled Water
268K, 0.34 MPa
(22.3 °F, 50 psia)
Chilled Recycled Water
118
Recycle Water
Chiller
274K (34°F)
2 07/18/05 Add SI units AJN
1 07/16/05 AJN
No. DATE REVISIONS BY CHKD DESIGN SUPV. ENGR PROJ ENGR APPRV
SCALE: DESIGNED: DRAWN:
NEXANT
San Francisco
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
ONCE-THROUGH CASE
AMMMONIA COMPRESSION REFRIGERATION
JOB No. DRAWING No. REV
00316 Figure 2 2
5 Dec, 2006 FILE: ONCE THRU NH3 REFRIG.VSD
page 1 of 1

2 .1 Utility Balance Ammonia Absorption Refrigeration

Item No	Item Name	Load BHP	Elect. Power	Steam (M Lbs per Hour)														Water, Mlbs/hr			C.W. circ.	Cooling, MMBTU / HR		Fuel MMBTU / HR	
		Norm.	KW	1950 psig	800 psig	650 psig	370 psig	330 psig	235 psig	50 psig	20 psig	20 psig SH	15 psig Sat	235 psig cond	50 psig cond	15 psig cond	2" HgA cond	BFW	Blow Down	Waste Water	gpm	Air	Water	Heat Abs.	Total Fired Duty
	Power Block																								
	HRSG HP Evaporator			(324.82)														327.23	(2.42)						
	Fired Tube Boiler			(754.61)														760.23	(5.61)						
	HTSC #1 Raw Gas Cooler (HP)			-														-	-						
	HP Steam Turbine (1,800 psig -> 370 psig)		50,284	981.17			(963.47)	(17.70)																	
	Gas Turbine Cooling						296.74	(296.74)																	
	Gas Turbine Injection						239.51																		
	Cold Reheat to HRSG						420.12	(420.12)																	
	PRV 1950 psig --> 250 psig		1,309	47.12					(47.12)																
	IP Steam Turbine (330 psig -> 55 psig)		39,697					736.43		(736.43)															
	LP Steam Turbine (43 psig -> 2 in HgA)		-							-			-												
	Makeup Condensate																(350.54)								
	Misc Steam Gen & Usages			2.97			5.35	(2.97)		(5.35)															
	Steam Seal Regulator Usage			0.61			1.76	1.10									(3.47)								
	LP Steam Turbine (43 psig -> 35 psia)		9,094							622.06		(622.06)													
	35 psia Stm desuperheat										(697.80)	622.06						75.74							
	LP HRSG Evap 20 psig stm										(44.93)							44.93							
	50 psig Steam Boiler										(47.89)							48.85	(0.96)					45.94	54.04
	Subtotal Power Block(Note 4)		100,384	(47.55)	-	-	0.00	0.00	(47.12)	(119.71)	(790.62)	-		-	-		(354.01)	1,256.97	(8.99)	-	-	-	-	45.94	54.04
	Gasification & Syngas Cleanup																								
	Candle Filter Raw Gas Cooler									-								-	-						
	HTSC #2 Raw Gas Cooler									-								-	-						
	LTSC Raw Gas Cooler									-								-	-						
	Injection to 50 psig Header									(19.32)								19.32							
	Claus Thermal Stage & WHB					(7.12)			(4.56)									11.68	(0.05)						
	HTSC #1 Raw Gas Cooler (MP)								-									-	-						
	Sulfur Condenser #1, 2 & 3									(5.69)								5.69	(0.04)						
	TGTU WHB									(1.62)								1.62	(0.01)						
	Letdown 235 psig --> 50 psig								-	-															
	Claus/TGTU Preheaters					7.12								(7.12)											
	Air Separation Plant								8.23					(8.23)											
	Sour Water Stripper								23.49					(23.49)											
	Oxygen Heater								5.62					(5.62)											
	Slurry Heater (MP)								29.38					(29.38)											
	CO2 Removal Plant									-					-						5,328		40.0		
	Claus TGTU Amine Regenerator									88.54					(88.54)										
	Waste Water Treatment									20.51					(20.51)										
	Process Steam									4.78					(4.78)										
	Slurry Heater (LP)									56.85					(56.85)										
	Shift Reactors				-													-							
	Low Temperature Cond Heater																								
	Stm to NH3 Absorp Reboiler										805.35					(805.35)									
	NH3 Absorption Regenerator Condenser																				22,740		113.7		
	NH3 Absorption Absorber Cooler																				132,660		663.3		
	Blowdown Flash										(3.17)								9.10	(5.92)					
	HP Cond Flash			47.55					(15.04)					(32.52)											
	MP Cond Flash									(24.34)				106.35	(82.01)										
	LP Cond Flash										(25.12)				170.68	(145.27)									
	MIscellaneous Cooling loads																				38,280		382.8		
	Cond to Fuel Gas Saturator														82.01	(782.42)									
	Syngas final cooler																				1,087		16.3		
	Subtotal Gasifier/Syngas Cleaning	-	-	47.55	-	-	-	-	47.12	119.71	777.05	-		-	-	(1,733.04)	-	38.30	8.99	(5.92)	200,095	-	1,216.1	-	-
	Deaerator																								
	Deaerator										13.57					927.69	354.01	(1,295.27)							
	TOTAL	-	100,384	-	-	-	0.00	0.00	0.00	0.00	0.00	-		-	-	-	-	0.00	-	(5.92)	200,095	-	1,216.1	45.94	54.04

1 All figures shown above represent normal utility usage requirements.
2 All figures shown above represent totals (duty, HP, KW, etc) unless indicated otherwise.
3 Cooling water supply temperature is 80 F. Makeup water to cooling tower is not shown.
4 Turbine electrical efficiency is back calculated from EPRI case 3A performance table power block kW output.
5 Steam Boiler Blowdown = 0.74 %
6 Sulfur production for EPRI 3B = 87 T/D
7 Delta cooling water requirements based on UOP supplied Selexol cooling water requirements of 16,193 gpm
8 Total Condensate to Fuel Gas Saturation is 23.24+82.01 = 105.25 Mlbs/hr.

No.	DATE		PROC.	PROJECT	CLIENT
Nexant			JOB No.		
			Drawing No.		REV.
			Sheet 2-1 Sheet 1		0

Item No	Item Name	Load BHP	Elect. Power	Steam (M Lbs per Hour)														Water, Mlbs/hr			C.W. circ.	Cooling, MMBTU / HR		Fuel MMBTU / HR	
		Norm.	KW	1950 psig	800 psig	650 psig	370 psig	330 psig	235 psig	50 psig	12 psig			235 psig cond	50 psig cond	12 psig cond	2" HgA cond	BFW	Blow Down	Waste Water	gpm	Air	Water	Heat Abs.	Total Fired Duty
	Gross Power Summary																								
	Gas Turbine Power		337,472																						
	Steam Turbine Power		100,384																						
	Generator Loss		(7,215)																						
	Turbo-Set Power		430,641																						
	Fuel Gas Expander Power		-																						
	Gross Plant Power		430,641																						
	Auxiliary Load Summary																								
	Coal Handling and Conveying		330																						
	Coal Milling		750																						
	Coal Slurry Pumps		200																						
	Slag Handling and Dewatering		150																						
	Scrubber Pumps		300																						
	Total Gasifier		1,730																						
	Recycle Gas Blower		600																						
	Air Separation Plant		23,330																						
	Oxygen Booster Compressor		11,910																						
	CO2 Removal Plant		9,465																						
	NH3 Absorption Refrigeration Plant		1,982																						
	MDEA Acid Gas Removal		1,300																						
	Claus/TGTU		100																						
	Tail Gas Recycle		1,410																						
	Humidification Tower Pump		100																						
	Humiditier Makeup Pump		60																						
	Low Pressure CO2 Compressor																								
	High Pressure CO2 Compressor																								
	Condensate Pumps		280																						
	High Pressure Boiler Feed Water Pump		2,940																						
	Low Pressure Boiler Feed Water Pump																								
	Misc. Balance of Plant		1,000																						
	Gas Turbine Auxiliaries		600																						
	Steam Turbine Auxiliaries		200																						
	Circulating Water Pumps		4,421																		200,095				
	Cooling Tower Fans		1,263																						
	Flash Bottoms Pump		50																						
	Transformer Loss		1,090																						
	Total Auxiliary Power Requirement		63,831																						
			54,435																						
	NET PLANT POWER		366,811																		200,095				

1 All figures shown above represent normal utility usage requirements.
2 All figures shown above represent totals (duty, HP, KW, etc) unless indicated otherwise.
3 Cooling water supply temperature is 80 F. Makeup water to cooling tower is not shown.
4 Turbine electrical efficiency is back calculated from EPRI case 3A performance table power block kW output.
5 Steam Boiler Blowdown = 0.60 %
6 Sulfur production for EPRI 3B = 87 T/D
7 Delta cooling water requirements based on UOP supplied Selexol cooling water requirements of 16,193 gpm
8 Total Condensate to Fuel Gas Saturation is 23.24+82.01 = 105.25 Mlbs/hr.

No.	DATE		PROC.	PROJECT	CLIENT
Nexant			**JOB No.**		
			Drawing No.		REV.
			Table 2-1 Sheet 2		0

Table 2.2 **Utility Balance Ammonia Compression Refrigeration**

Item No	Item Name	Load BHP	Elect. Power	Steam (M Lbs per Hour)													Water, Mlbs/hr			C.W. circ.	Cooling, MMBTU / HR		Fuel MMBTU / HR	
		Norm.	KW	1950 psig	800 psig	650 psig	370 psig	330 psig	235 psig	50 psig	12 psig		235 psig cond	50 psig cond	12 psig cond	2" HgA cond	BFW	Blow Down	Waste Water	gpm	Air	Water	Heat Abs.	Total Fired Duty
	Power Block																							
	HRSG HP Evaporator			(324.82)													327.23	(2.42)						
	Fired Tube Boiler			(754.61)													760.23	(5.61)						
	HTSC #1 Raw Gas Cooler (HP)			-													-	-						
	HP Steam Turbine (1,800 psig -> 370 psig)		50,284	981.17			(963.47)	(17.70)																
	Gas Turbine Cooling						296.74	(296.74)																
	Gas Turbine Injection						239.51																	
	Cold Reheat to HRSG						420.12	(420.12)																
	PRV 1950 psig --> 250 psig		1,309	47.12					(47.12)															
	IP Steam Turbine (330 psig -> 55 psig)		39,697					736.43		(736.43)														
	LP Steam Turbine (43 psig -> 2 in HgA)		52,051							622.06						(622.06)								
	Makeup Condensate															(349.59)								
	Misc Steam Gen & Usages			2.97			5.35	(2.97)		(5.35)														
	Steam Seal Regulator Usage			0.61			1.76	1.10								(3.47)								
	Surface Condenser																			121,720		608.60		
	Subtotal Power Block(Note 4)		143,341	(47.55)	-	-	0.00	0.00	(47.12)	(119.71)	-	-	-	-		(975.12)	1,087.46	(8.03)	-	121,720	-	608.60	-	-
	Gasification & Syngas Cleanup																							
	Candle Filter Raw Gas Cooler									-							-	-						
	HTSC #2 Raw Gas Cooler									-							-	-						
	LTSC Raw Gas Cooler									-							-	-						
	Injection to 50 psig Header									(19.32)							19.32							
	Claus Thermal Stage & WHB					(7.12)			(4.56)								11.68	(0.05)						
	HTSC #1 Raw Gas Cooler (MP)								-								-	-						
	Sulfur Condenser #1, 2 & 3									(5.69)							5.69	(0.04)						
	TGTU WHB									(1.62)							1.62	(0.01)						
	Letdown 235 psig --> 50 psig								-	-														
	Claus/TGTU Preheaters					7.12							(7.12)											
	Air Separation Plant								8.23				(8.23)											
	Sour Water Stripper								23.49				(23.49)											
	Oxygen Heater								5.62				(5.62)											
	Slurry Heater (MP)								29.38				(29.38)											
	CO2 Removal Plant																			5,328		40.0		
	Ammonia Compression Refrigeration									-				-						10,272		77.0		
	Claus TGTU Amine Regenerator									88.54				(88.54)										
	Waste Water Treatment									20.51				(20.51)										
	Process Steam									4.78				(4.78)										
	Slurry Heater (LP)									56.85				(56.85)										
	Shift Reactors				-												-							
	Low Temperature Cond Heater																							
	Blowdown Flash										(3.17)							8.14	(4.97)					
	HP Cond Flash			47.55					(15.04)				(32.52)											
	MP Cond Flash									(24.34)			106.35	(82.01)										
	LP Cond Flash										(27.12)			170.68	(143.58)									
	Cond to Fuel Gas Saturator													82.01	23.24									
	Miscelaneous cooling Loads																			38,280		383		
	Subtotal Gasifier/Syngas Cleaning	-	-	47.55	-	-	-	-	47.12	119.71	(30.30)	-	-	-	(120.34)	-	38.30	8.03	(4.97)	53,880	-	500	-	-
	Deaerator																							
	Deaerator										30.30				120.34	975.12	(1,125.76)							
	TOTAL	-	143,341	-	-	-	0.00	0.00	0.00	0.00	-	-	-	-	-	-	(0.00)	-	(4.97)	175,600	-	1,108.40	-	-

1 All figures shown above represent normal utility usage requirements.
2 All figures shown above represent totals (duty, HP, KW, etc) unless indicated otherwise.
3 Cooling water supply temperature is 80 F. Makeup water to cooling tower is not shown.
4 Turbine electrical efficiency is back calculated from EPRI case 3A performance table power block kW output.
5 Steam Boiler Blowdown = 0.74 %
6 Sulfur production for EPRI 3B = 87 T/D
7 Delta cooling water requirements based on UOP supplied Selexol cooling water requirements of 16,193 gpm
8 Total Condensate to Fuel Gas Saturation is 23.24+82.01 = 105.25 Mlbs/hr.

No.	DATE		PROC.	PROJECT	CLIENT
Nexant			JOB No.		
			Drawing No.		REV.
			Table 2.2 Sheet 1		0

Item No	Item Name	Load BHP	Elect. Power	Steam (M Lbs per Hour)													Water, Mlbs/hr			C.W. circ.	Cooling, MMBTU / HR		Fuel MMBTU / HR	
		Norm.	KW	1950 psig	800 psig	650 psig	370 psig	330 psig	235 psig	50 psig	12 psig		235 psig cond	50 psig cond	12 psig cond	2" HgA cond	BFW	Blow Down	Waste Water	gpm	Air	Water	Heat Abs.	Total Fired Duty
	Gross Power Summary																							
	Gas Turbine Power		337,472																					
	Steam Turbine Power		143,341																					
	Generator Loss		(7,215)																					
	Turbo-Set Power		473,598																					
	Fuel Gas Expander Power		-																					
	Gross Plant Power		473,598																					
	Auxiliary Load Summary																							
	Coal Handling and Conveying		330																					
	Coal Milling		750																					
	Coal Slurry Pumps		200																					
	Slag Handling and Dewatering		150																					
	Scrubber Pumps		300																					
	Total Gasifier		1,730																					
	Recycle Gas Blower		600																					
	Air Separation Plant		23,330																					
	Oxygen Booster Compressor		11,910																					
	CO2 Removal Plant		9,465																					
	NH3 Compression Refrigeration		19,958																					
	MDEA Acid Gas Removal		1,300																					
	Claus/TGTU		100																					
	Tail Gas Recycle		1,410																					
	Humidification Tower Pump		100																					
	Humiditier Makeup Pump		60																					
	Low Pressure CO2 Compressor																							
	High Pressure CO2 Compressor																							
	Condensate Pumps		280																					
	High Pressure Boiler Feed Water Pump		2,940																					
	Low Pressure Boiler Feed Water Pump																							
	Misc. Balance of Plant		1,000																					
	Gas Turbine Auxiliaries		600																					
	Steam Turbine Auxiliaries		200																					
	Circulating Water Pumps		3,879																	175,600				
	Cooling Tower Fans		1,108																					
	Flash Bottoms Pump		50																					
	Transformer Loss		1,090																					
	Total Auxiliary Power Requirement		81,111																					
			54,435																					
	NET PLANT POWER		392,487																	175,600				

1 All figures shown above represent normal utility usage requirements.
2 All figures shown above represent totals (duty, HP, KW, etc) unless indicated otherwise.
3 Cooling water supply temperature is 80 F. Makeup water to cooling tower is not shown.
4 Turbine electrical efficiency is back calculated from EPRI case 3A performance table power block kW output.
5 Steam Boiler Blowdown = 0.60 %
6 Sulfur production for EPRI 3B = 87 T/D
7 Delta cooling water requirements based on UOP supplied Selexol cooling water requirements of 16,193 gpm
8 Total Condensate to Fuel Gas Saturation is 23.24+82.01 = 105.25 Mlbs/hr.

No.	DATE		PROC.	PROJECT	CLIENT

Nexant

JOB No.
Drawing No. REV.
Table 2-2 Sheet 2

Task 2 - CO_2 Capture with Promoted Solvent

Objective

Task 2 investigates the effect of a promoter (e.g., tetrahydrofuran or THF) on the Simteche CO_2 capture process using a one-stage hydrate reactor. It has been observed at LANL that a promoter can significantly alter the phase envelope of CO_2 hydrate formation, by increasing the HFT (maximum hydrate formation temperature) at which CO_2 hydrate formation can occur at a given CO_2 partial pressure to a higher temperature. This would allow the Simteche process to be operated at higher temperatures, which may result in potential savings in refrigeration cost. In a previous study (Phase II Engineering Analysis), the effect of promoter was investigated in a two-stage process.

The objectives of the current task are to (1) develop a single-stage promoter process design, complete with heat, material and utility balances, and capital cost estimates to allow its techno-economic feasibility to be assessed, and (2) compare the CO_2 capture cost of this process with the two-stage promoter process and with the Selexol process presented in the February 2006 Phase II Engineering Analysis Report Final Version.

Summary

The following design basis was used in developing the one-stage promoter hydrate process using tetrahydrofuran (THF) as the promoter, since its laboratory result represents the best of all the promoters tested thus far at Los Alamos National Laboratory (LANL)[1] These design basis are:

- LANL's experimental investigation of THF as a promoter forming a 'SII structure 1' type of clathrate with the following property –
    - Hydration number of 8.5
    - Heat of hydrate formation of 89.20 kJ/gmol CO_2
- Shifted syngas feed to the battery limit of the hydrate formation reactor at 1,010 psia and 105 ^oF. The feed syngas is cooled to 43°F before entering the hydrate reactor,
- Hydrate formation reactor operates at 43°F and 1,000 psia,
- Hydrate slurry separator operates at 43° and 940 psia,
- High pressure hydrate flash reactor operates at 70°F and 625 psia,
- Low pressure hydrate flash reactor operates at 70°F and 285 psia, and
- Ammonia vapor compression refrigeration is used to provide process cooling duty requirements.

The heat and material balance was developed for this case based on the assumption that 90% CO_2 capture is achievable with the use of a promoter solvent. This

[1] R. Currier et. al., 5th Annual Conference on Carbon Capture & Sequestration, May 2006

assumption needs to be verified once the promoter hydrate equilibrium correlations are developed based on the work from LANL. The promoter hydrate equilibrium development is still at the experimental stage, and the preliminary data suggests that 90% CO_2 capture can be achieved.

Additionally, a hydration number of 8.5 used for this analysis also needs to be verified based on further LANL promoter work. During the course of this study, it was noted that this hydration number might be optimistic based on the latest data from LANL, which suggests that the hydration number should be higher (approximately 12). High hydration number would increase the reactor cooling and refrigeration duty, thereby increasing operating and capital costs.

The block flow diagram for the one-stage promoter hydrate reactor is shown in Figures 1-1 and 1-2.

Methodology

The material balance spreadsheet model created for the Phase II Engineering Analysis study was used to develop the heat and material balance for the hydrate process area and HYSYS (a commercial process simulator) was used to simulate the ammonia refrigeration processing loop. The material balance was used to estimate equipment sizing, utilities requirements, and refrigeration load for the process. ICARUS was used to develop the bare equipment cost estimates. An installation factor appropriate to the type of equipment use (i.e., 1.7 for compressors and 3.4 for other process equipments) was used to obtain the total installed cost for the process. Similar installation factors were used in the February 2006 Report.

Results

The results are presented in the following:

Table 1-1 Material Balance
Table 1-2 Utilities Summary
Table 1-3 Equipment List
Table 1-4 Equipment Cost Estimate
Table 1-5 Power Production Summary
Table 1-6 Carbon Control Costs

The total installed cost for the one-stage promoter process, expressed in 2nd quarter 2005 dollars, is $ 100.1 million.

Table 1-4 shows the total equipment cost comparison of the one-stage promoter process with the two-stage promoter process and the Selexol process reported in 2005. As can been seen, the cost of the one-stage promoter process is significantly more expensive. The reason for that is mainly due to the increase in the heat of hydrate formation associated with a SII clathrate structure promoter. The almost 44% increase in hydrate heat of formation (from 62.0 to 89.2 kJ/gmole) resulted in significant increase in refrigeration duty, hence larger equipment sizes and costs. In addition, a considerable increase in refrigeration compression horsepower is also required for the ammonia refrigeration system. The higher power consumption resulted in a lower net power production and plant efficiency, when compared with the two-stage promoter process and the Selexol process.

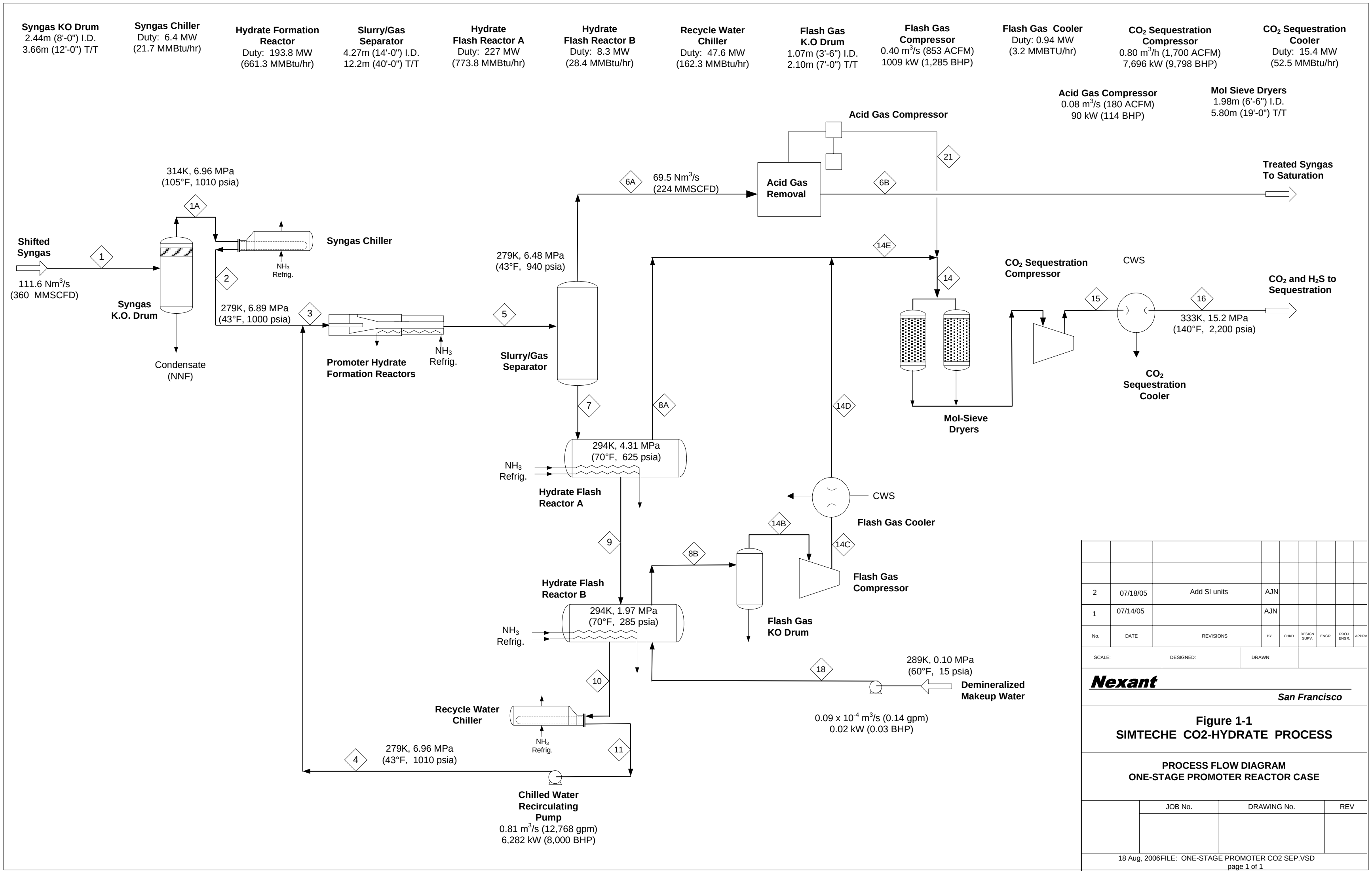
Table 1-5 shows the overall IGCC plant production summary. As shown, the parasitic load of the one-stage promoter process represents approximately 12% of plant gross power output compared to 7.5% and 9% for the Selexol and two-stage promoter processes, respectively.

Table 1-6 shows the CO_2 captured cost for three processes. The net capture cost of \$19/ton of CO_2 capture for the one-stage promoter process is approximately 27% to 58% higher than that of the Selexol and the two-stage promoter processes.

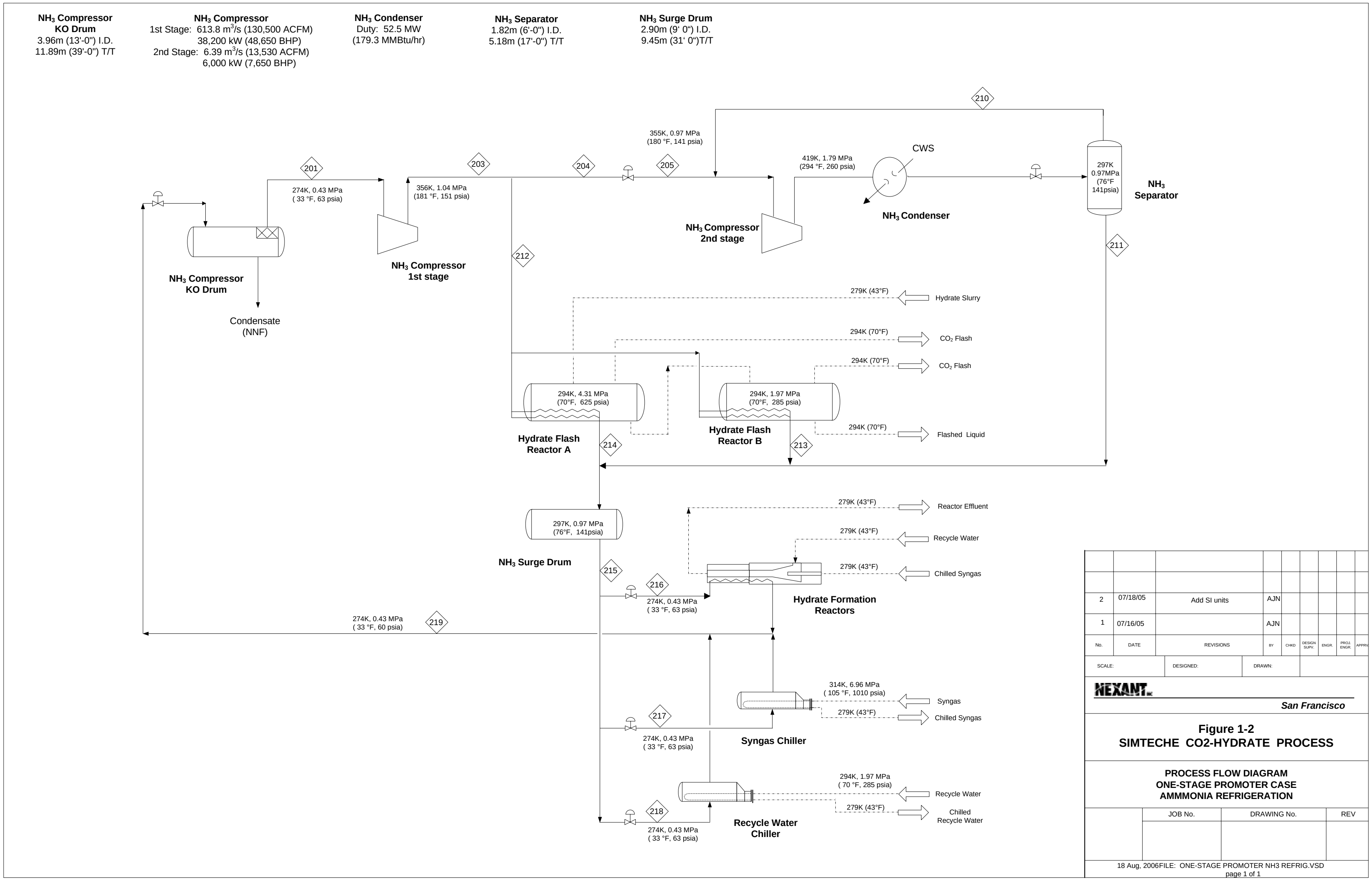
The cost of electricity of 6.56 cents per kWhr of the one-stage promoter process represents an increase of approximately 43% compared to the No-Capture Case.

Conclusion

While the addition of a promoter additive allows the Simteche CO_2 capture process to be operated at a higher temperature, the CO_2 hydrate formed has a different clathrate structure with a higher hydration number and heat of formation. As a result, significantly higher refrigeration loading and ammonia refrigeration compression horsepower are required for the subject one-stage promoter process, making it even less competitive when compared to the two-stage promoter process and the Selexol process studied earlier.



Syngas KO Drum
2.44m (8'-0") I.D.
3.66m (12'-0") T/T
Syngas Chiller
Duty: 6.4 MW
(21.7 MMBtu/hr)
Hydrate Formation Reactor
Duty: 193.8 MW
(661.3 MMBtu/hr)
Slurry/Gas Separator
4.27m (14'-0") I.D.
12.2m (40'-0") T/T
Hydrate Flash Reactor A
Duty: 227 MW
(773.8 MMBtu/hr)
Hydrate Flash Reactor B
Duty: 8.3 MW
(28.4 MMBtu/hr)
Recycle Water Chiller
Duty: 47.6 MW
(162.3 MMBtu/hr)
Flash Gas K.O Drum
1.07m (3'-6") I.D.
2.10m (7'-0") T/T
Flash Gas Compressor
0.40 m³/s (853 ACFM)
1009 kW (1,285 BHP)
Flash Gas Cooler
Duty: 0.94 MW
(3.2 MMBTU/hr)
CO2 Sequestration Compressor
0.80 m³/h (1,700 ACFM)
7,696 kW (9,798 BHP)
CO2 Sequestration Cooler
Duty: 15.4 MW
(52.5 MMBtu/hr)
Acid Gas Compressor
0.08 m³/s (180 ACFM)
90 kW (114 BHP)
Mol Sieve Dryers
1.98m (6'-6") I.D.
5.80m (19'-0") T/T
Acid Gas Compressor
314K, 6.96 MPa
(105°F, 1010 psia)
1A
6A
69.5 Nm³/s
(224 MMSCFD)
Acid Gas Removal
6B
21
Treated Syngas To Saturation
Shifted Syngas
1
111.6 Nm³/s
(360 MMSCFD)
2
Syngas Chiller
NH3 Refrig.
279K, 6.48 MPa
(43°F, 940 psia)
14E
14
CO2 Sequestration Compressor
CWS
CO2 and H2S to Sequestration
15
16
333K, 15.2 MPa
(140°F, 2,200 psia)
Syngas K.O. Drum
279K, 6.89 MPa
(43°F, 1000 psia)
3
5
Condensate
(NNF)
Promoter Hydrate Formation Reactors
NH3 Refrig.
Slurry/Gas Separator
CO2 Sequestration Cooler
7
8A
14D
Mol-Sieve Dryers
294K, 4.31 MPa
(70°F, 625 psia)
NH3 Refrig.
Hydrate Flash Reactor A
CWS
Flash Gas Cooler
14B
9
8B
14C
Flash Gas Compressor
Hydrate Flash Reactor B
294K, 1.97 MPa
(70°F, 285 psia)
NH3 Refrig.
Flash Gas KO Drum
18
289K, 0.10 MPa
(60°F, 15 psia)
Demineralized Makeup Water
10
Recycle Water Chiller
0.09 x 10⁻⁴ m³/s (0.14 gpm)
0.02 kW (0.03 BHP)
NH3 Refrig.
11
4
279K, 6.96 MPa
(43°F, 1010 psia)
Chilled Water Recirculating Pump
0.81 m³/s (12,768 gpm)
6,282 kW (8,000 BHP)
2 07/18/05 Add SI units AJN
1 07/14/05 AJN
No. DATE REVISIONS BY CHKD DESIGN SUPV. ENGR. PROJ. ENGR. APPRV.
SCALE: DESIGNED: DRAWN:
Nexant
San Francisco
Figure 1-1
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
ONE-STAGE PROMOTER REACTOR CASE
JOB No. DRAWING No. REV
18 Aug, 2006FILE: ONE-STAGE PROMOTER CO2 SEP.VSD
page 1 of 1



NH3 Compressor KO Drum
3.96m (13'-0") I.D.
11.89m (39'-0") T/T
NH3 Compressor
1st Stage: 613.8 m3/s (130,500 ACFM)
38,200 kW (48,650 BHP)
2nd Stage: 6.39 m3/s (13,530 ACFM)
6,000 kW (7,650 BHP)
NH3 Condenser
Duty: 52.5 MW
(179.3 MMBtu/hr)
NH3 Separator
1.82m (6'-0") I.D.
5.18m (17'-0") T/T
NH3 Surge Drum
2.90m (9' 0") I.D.
9.45m (31' 0")T/T
210
355K, 0.97 MPa
(180 °F, 141 psia)
201
203
204
205
419K, 1.79 MPa
(294 °F, 260 psia)
CWS
274K, 0.43 MPa
(33 °F, 63 psia)
356K, 1.04 MPa
(181 °F, 151 psia)
297K
0.97MPa
(76°F
141psia)
NH3 Separator
NH3 Compressor 2nd stage
NH3 Condenser
211
212
NH3 Compressor 1st stage
NH3 Compressor KO Drum
Condensate (NNF)
279K (43°F)
Hydrate Slurry
294K (70°F)
CO2 Flash
294K (70°F)
CO2 Flash
294K, 4.31 MPa
(70°F, 625 psia)
294K, 1.97 MPa
(70°F, 285 psia)
294K (70°F)
Flashed Liquid
Hydrate Flash Reactor A
214
Hydrate Flash Reactor B
213
279K (43°F)
Reactor Effluent
297K, 0.97 MPa
(76°F, 141psia)
279K (43°F)
Recycle Water
NH3 Surge Drum
279K (43°F)
Chilled Syngas
215
216
274K, 0.43 MPa
(33 °F, 63 psia)
Hydrate Formation Reactors
274K, 0.43 MPa
(33 °F, 60 psia)
219
314K, 6.96 MPa
(105 °F, 1010 psia)
Syngas
217
279K (43°F)
Chilled Syngas
274K, 0.43 MPa
(33 °F, 63 psia)
Syngas Chiller
294K, 1.97 MPa
(70 °F, 285 psia)
Recycle Water
218
279K (43°F)
Chilled Recycle Water
274K, 0.43 MPa
(33 °F, 63 psia)
Recycle Water Chiller
2 07/18/05 Add SI units AJN
1 07/16/05 AJN
No. DATE REVISIONS BY CHK'D DESIGN SUPV. ENGR. PROJ. ENGR. APPRV.
SCALE: DESIGNED: DRAWN:
NEXANT INC
San Francisco
Figure 1-2
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
ONE-STAGE PROMOTER CASE
AMMMONIA REFRIGERATION
JOB No. DRAWING No. REV
18 Aug, 2006FILE: ONE-STAGE PROMOTER NH3 REFRIG.VSD
page 1 of 1

Table 1-1 Material Balance

Stream Numbers		Strm 1	Strm 1	Strm 1A	Strm 1A	Strm 1B	Strm 1B	Strm 2	Strm 2	Strm 3	Strm 3	Strm 4	Strm 4	Strm 5	Strm 5
		Sat'd Syngas to KO drum	Sat'd Syngas to KO drum	Sat'd Syngas to chiller	Sat'd Syngas to chiller	KO drum Condensate	KO drum Condensate	Chilled Syngas	Chilled Syngas	Hform Rx Feed	Hform Rx Feed	Chilled Recy Water to 1st Stg Rx	Chilled Recy Water to 1st Stg Rx	Reactor Effluent	Reactor Effluent
	Mol Wt	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h
H_2	2.02	21,418.7	43,175.9	21,418.7	43,175.9	-	-	21,418.7	43,175.9	21,418.8	43,175.9	0.0	0.1	21,418.8	43,175.9
CO_2	44.01	16,381.1	720,930.3	16,381.1	720,930.3	-	-	16,381.1	720,930.3	20,289.0	892,914.2	3,907.9	171,983.9	2,183.4	96,089.4
H_2S	34.08	248.4	8,465.8	248.4	8,465.8	-	-	248.4	8,465.8	330.7	11,268.4	82.2	2,802.6	330.7	11,268.4
H_2O Vap	18.02	4.0	72.8	4.0	72.8	-	-	4.0	72.8	4.0	72.8	-	-	4.0	72.8
Ar	39.95	292.7	11,693.6	292.7	11,693.6	-	-	292.7	11,693.6	292.7	11,693.6	0.0	0.0	292.7	11,693.6
N2	28.01	289.3	8,104.8	289.3	8,104.8	-	-	289.3	8,104.8	289.3	8,104.8	0.0	0.0	289.3	8,104.8
CO	28.01	382.3	10,708.5	382.3	10,708.5	-	-	382.3	10,708.5	382.3	10,708.6	0.0	0.1	382.3	10,708.6
CH4	16.04	537.8	8,627.8	537.8	8,627.8	-	-	537.8	8,627.8	537.8	8,627.9	0.0	0.1	537.8	8,627.9
COS	60.07	0.0	2.4	0.0	2.4	-	-	0.0	2.4	0.1	3.1	0.0	0.7	0.1	3.1
NH3	17.03	3.4	57.9	3.4	57.9	-	-	3.4	57.9	303.4	5,166.4	300.0	5,108.5	303.4	5,166.4
H_2O Liq	18.02	-	-	-	-	-	-	-	-	278,552.1	5,018,172.5	278,552.1	5,018,172.5	124,654.4	2,245,673.3
CO_2 Hydrate	152.10	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2S Hydrate	142.17	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Promoter	72.00	-								16,461.7	1,185,243.8	16,461.7	1,185,243.8	7,408.9	533,441.5
CO2/Pro Hydrate	3,730.22	-								-		-		1,131.6	4,221,126.3
Total		39,557.91	811,840	39,558	811,840	-	-	39,558	811,840	338,862	7,195,152	299,304	6,383,312	158,937	7,195,152
Temperature, F		105		105		105		43		43		43		43	
Pressure, psia		1,010		1,010		1,010		940		1,000		1,000		920	

Stream Numbers		Strm 6A	Strm 6A	Strm 6B	Strm 6B	Strm 7	Strm 7	Strm 8A	Strm 8A	Strm 8B	Strm 8B	Strm 9	Strm 9	Strm 10	Strm 10
		Treated SG to AGR	Treated SG to AGR	Treated SG to GT	Treated SG to GT	Slurry to HFR-A	Slurry to HFR-A	CO_2 Vapor fr HFR-A	CO_2 Vapor fr HFR-A	CO_2 Vapor fr HFR-B	CO_2 Vapor fr HFR-B	Water fr fr HFR-A	Water fr fr HFR-A	Water fr fr HFR-B	Water fr fr HFR-B
	Mol Wt	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h
H_2	2.02	21,302.8	42,942.3	21,302.8	42,942.3	115.9	233.7	114.3	230.3	1.6	3.3	1.7	3.4	0.0	0.1
CO_2	44.01	1,638.2	72,096.0	1,605.9	70,673.6	545.2	23,993.4	11,854.9	521,731.7	2,888.1	127,102.6	6,795.9	299,086.5	3,907.9	171,983.9
H_2S	34.08	170.2	5,799.2	148.3	5,054.0	160.5	5,469.1	58.3	1,988.2	19.9	678.4	102.2	3,481.0	82.2	2,802.6
H_2O Vap	18.02	3.3	60.0	31.6	569.6	-	-	6.4	115.9	3.4	62.1	-	-	-	-
Ar	39.95	291.2	11,633.3	291.2	11,633.3	1.5	60.3	1.5	59.6	0.0	0.7	0.0	0.7	0.0	0.0
N2	28.01	287.8	8,063.1	287.8	8,063.1	1.5	41.8	1.5	41.3	0.0	0.5	0.0	0.5	0.0	0.0
CO	28.01	379.3	10,625.4	379.3	10,625.4	3.0	83.2	2.9	81.5	0.1	1.6	0.1	1.7	0.0	0.1
CH4	16.04	531.6	8,528.5	531.6	8,528.5	6.2	99.4	6.1	97.1	0.1	2.2	0.1	2.3	0.0	0.1
COS	60.07	0.0	1.6	0.0	1.6	0.0	1.5	0.0	0.5	0.0	0.2	0.0	0.9	0.0	0.7
NH3	17.03	2.0	34.9	2.0	34.9	301.3	5,131.5	0.9	15.0	0.5	8.0	300.4	5,116.5	300.0	5,108.5
H_2O Liq	18.02	-	-	-	-	124,655.1	2,245,686.0	-	-	-	-	278,546.4	5,018,069.4	278,552.1	5,018,172.5
CO_2 Hydrate	152.10	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2S Hydrate	142.17	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Promoter	72.00	2.2	157.2	2.2	157.2	7,406.7	533,284.3	3.6	257.0	1.9	137.7	16,456.0	1,184,829.6	16,461.7	1,185,243.8
CO2/Pro Hydrate	3,730.22	-	-	-	-	1,131.6	4,221,126.3	-	-	-	-	-	-	-	-
Total		24,609	159,941	24,582.87	158,283	134,329	7,035,211	12,050	524,618	2,916	127,997	302,203	6,510,592	299,304	6,383,312
Temperature, F		43		105		43		70		70		70		70	
Pressure, psia		940		852		940		625		285		625		285	

Table 1-1 Material Balance, cont'd

Stream Numbers		Strm 11	Strm 11	Strm 14A	Strm 14A	Strm 14B	Strm 14B	Strm 14C	Strm 14C	Strm 14D	Strm 14D	Strm 14E	Strm 14E	Strm 14	Strm 14
		Chilled Recycle Water	Chilled Recycle Water	FlashGas Compr KO Liq	FlashGas Compr KO Liq	Flash Gas Compr Feed (HFR-B)	Flash Gas Compr Feed (HFR-B)	Flash Gas Compr Discharge	Flash Gas Compr Discharge	Cooled Flash Gas (HFR-B)	Cooled Flash Gas (HFR-B)	Mixed CO_2 to Dryer	Mixed CO_2 to Dryer	Total CO2 to Drying	Total CO2 to Drying
	Mol Wt	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h
H_2	2.02	0.0	0.1	-	-	1.6	3.3	1.6	3.3	1.6	3.3	115.9	233.6	115.9	233.6
CO_2	44.01	3,907.9	171,983.9	-	-	2,888.1	127,102.6	2,888.1	127,102.6	2,888.1	127,102.6	14,743.0	648,834.3	14,775.3	650,256.7
H_2S	34.08	82.2	2,802.6	-	-	19.9	678.4	19.9	678.4	19.9	678.4	78.3	2,666.6	100.1	3,411.8
H_2O Vap	18.02	-	-	-	-	3.4	62.1	3.4	62.1	3.4	62.1	9.9	178.0	10.1	182.5
Ar	39.95	0.0	0.0	-	-	0.0	0.7	0.0	0.7	0.0	0.7	1.5	60.3	1.5	60.3
N2	28.01	0.0	0.0	-	-	0.0	0.5	0.0	0.5	0.0	0.5	1.5	41.8	1.5	41.8
CO	28.01	0.0	0.1	-	-	0.1	1.6	0.1	1.6	0.1	1.6	3.0	83.1	3.0	83.1
CH4	16.04	0.0	0.1	-	-	0.1	2.2	0.1	2.2	0.1	2.2	6.2	99.3	6.2	99.3
COS	60.07	0.0	0.7	-	-	0.0	0.2	0.0	0.2	0.0	0.2	0.0	0.8	0.0	0.8
NH3	17.03	300.0	5,108.5	-	-	0.5	8.0	0.5	8.0	0.5	8.0	1.4	23.0	1.4	23.0
H_2O Liq	18.02	278,552.1	5,018,172.5	-	-	-	-	-	-	-	-	-	-	-	-
CO_2 Hydrate	152.10	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2S Hydrate	142.17	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Promoter	72.00	16,461.7	1,185,243.8	-	-	1.9	137.7	1.9	137.7	1.9	137.7	5.5	394.8	5.5	394.8
CO2/Pro Hydrate	3,730.22	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total		299,304	6,383,312	-	-	2,916	127,997	2,916	127,997	2,916	127,997	14,966	652,615	15,020	654,788
Temperature, F		43		70		70		204		110		77		78	
Pressure, psia		280		285		285		630		625		625		625	

Stream Numbers		Strm 15	Strm 15	Strm 16	Strm 16	Strm 17	Strm 17	Strm 18A	Strm 18A	Strm 18B	Strm 18B	Strm 21	Strm 21	Strm 21A	Strm 21A
		CO2 Seq Compr Discharge	CO2 Seq Compr Discharge	Cooled CO_2 Vapor to Seq	Cooled CO_2 Vapor to Seq	Total Condensate	Total Condensate	Demin Water Makeup	Demin Water Makeup	Promoter Makeup	Promoter Makeup	Acid Gas fr AGR	Acid Gas fr AGR	Acid Gas Compr KO Liq	Acid Gas Compr KO Liq
	Mol Wt	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h
H_2	2.02	115.9	233.6	115.9	233.6	-	-	-	-	-	-	-	-	-	-
CO_2	44.01	14,775.3	650,256.7	14,775.3	650,256.7	-	-	-	-	-	-	32.3	1,422.4	-	-
H_2S	34.08	100.1	3,411.8	100.1	3,411.8	-	-	-	-	-	-	21.9	745.2	-	-
H_2O Vap	18.02	-	-	-	-	-	-	-	-	-	-	0.3	4.5	-	-
Ar	39.95	1.5	60.3	1.5	60.3	-	-	-	-	-	-	-	-	-	-
N2	28.01	1.5	41.8	1.5	41.8	-	-	-	-	-	-	-	-	-	-
CO	28.01	3.0	83.1	3.0	83.1	-	-	-	-	-	-	-	-	-	-
CH4	16.04	6.2	99.3	6.2	99.3	-	-	-	-	-	-	-	-	-	-
COS	60.07	0.0	0.8	0.0	0.8	-	-	-	-	-	-	-	-	-	-
NH3	17.03	1.4	23.0	1.4	23.0	-	-	-	-	-	-	-	-	-	-
H_2O Liq	18.02	-	-	-	-	10.1	182.5	9.2	165.2	9.2	165.2	-	-	8.3	148.6
CO_2 Hydrate	152.10	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2S Hydrate	142.17	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Promoter	72.00	5.5	394.8	5.5	394.8	-	-	-	-	7.7	552.0	-	-	-	-
CO2/Pro Hydrate	3,730.22	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total		15,010	654,605	15,010	654,605	10	182	9	165	17	717	54	2,172	8	149
Temperature, F		78		140		100		60		60		246		110	
Pressure, psia		2,210		2,200		625		295		295		625		18	

Table 1-2 Utilities Summary

Item No	Item Name	Load BHP		Elect. Power	Cooling Water		Refrigeration	
		Norm.	Max (3).	KW	CW, MMbtu/hr	C.W. circ. GPM (2)	NH3, MMbtu/hr	Tons of Refrig.
	Exchangers							
	Syngas Chiller						21.5	1,795
	Hydrate Flash Reactor A							
	Hydrate Flash Reactor B							
	Recycle Water Chiller						160.3	13,358
	Flash Gas Cooler				3.2	427		
	CO_2 Sequestration Cooler				52.5	7,000		
	Acid Gas Compr 1st Intercooler				0.1	13		
	Acid Gas Compr 2nd Intercooler				0.1	13		
	Acid Gas Compr 3rd Intercooler				0.1	13		
	Ammonia Condenser				179.3	23,907		
	Mol Sieve Regeneration Heater							
	Hydrate Formation Reactors							
	Hydrate Formation Reactor						661.3	55,111
	Compressors							
	Flash Gas Compressor	1,285		1,009	0.07	9		
	CO_2 Sequestration Compressor	9,800		7,696	0.50	67		
	Acid Gas Compressor	115		90	0.01	1		
	NH_3 Compressor - Stage 1	48,650		38,203	2.48	330		
	NH_3 Compressor - Stage 2	7,650		6,007	0.39	52		
	Pumps							
	Chilled Water Recycle Pumps	8,000		6,282				
	Cooling Water Pumps	794		623				
	Lube Oil Pumps	22		17				
	Demineralized Water Pumps	0.03		0.0				
	TOTAL	76,316		59,928	238.74	31,831	843.2	70,264

NOTES:
1 All Figures shown above represent normal utility usage requirements except:
() indicates normal utility make
* indicates intermittent usage or make, not included in totals
2 Cooling water supply temperature is 80 F. Makeup water to cooling tower is not shown
3 Utility consumption for max. load conditions is not shown.

Table 1-3 Equipment List

Equipment	No. ea.	Size/Duty	Material Tube/Shell	Temperature (°F) Tube/Shell	Pressure (psig) Tube/Shell
Reactors					
Hydrate Formation Reactors	10	661.3 MM Btu/hr total	304 SS/CS	200/200	1100/235
Hydrate Flash Reactor A	6	773.8 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/680
Hydrate Flash Reactor B	1	28.4 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/315
Exchangers					
Syngas Chiller	1	21.7 MM Btu/hr	304 SS/CS	200/200	1,100/235
Recycle Water Chiller	2	162. MM Btu/hr	304 SS/CS	200/200	300/235
Flash Gas Cooler	1	3.2 MM Btu/hr	CS/CS	285/200	680/50
CO_2 Sequestration Cooler	1	52.5 MM Btu/hr	CS/CS	375/200	2,400/50
Ammonia Condenser	2	179.3 MM Btu/hr	CS/CS	200/360	500/235
Acid Gas Compressor 1st Stage Intercooler	1	0.1 MM Btu/hr	CS/304 SS (Clad)	200/350	50/50
Acid Gas Compressor 2nd Stage Intercooler	1	0.1 MM Btu/hr	CS/304 SS (Clad)	200/350	50/110
Acid Gas Compressor 3rd Stage Intercooler	1	0.1 MM Btu/hr	CS/304 SS (Clad)	200/350	50/320
Regeneration Heater-Electric	1	200 kW	CS/CS	400	15
Vessels					
Syngas KO Drum	1	8'ID x 12' T-T	304 SS Clad	155	1,055
Slurry/Gas Separator	1	14ID x 40' T-T	304 SS Clad	100	985
Flash Gas Compr KO Drum	1	3.5'ID x 7' T-T	304 SS Clad	120	300
Acid Gas Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	50
Acid Gas 1st Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	100
Acid Gas 2nd Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	230
Acid Gas Product KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	670
Mol Sieve Dryer	2	6'ID x 12' T-T	304 SS Clad	450	670
Ammonia Separator	1	6'ID x 17' T-T	CS	200	235
Ammonia Surge Drum	1	9.5'ID x 31' T-T	CS	200	235
Ammonia Compr KO Drum	2	13'ID x 39' T-T	CS	200	235
Compressors					
Flash Gas Compressor	1	1,415 HP	CS	350	625
CO2 Sequestration Compressor	1	10,800 HP	CS	350	2400
Acid Gas Compressor	1	125 HP	CS	350	610
NH3 Refrig 1st Stage Compressor	1	53,500 HP	CS	200	136
NH3 Refrig 2nd Stage Compressor	1	7,650 HP	CS	350	245
Pumps					
Chilled Water Recycle Pumps (3-50% pumps)	3	6,400 GPM/ 4,000 HP each	304 SS	200	1100
Cooling Water Pumps (3-50% pumps)	3	15,000 GPM/400 HP each	CS	200	100
Demineralized Water Pumps (2-100% pumps)	2	0.14 GPM/0.03 HP each	CS		
Lube Oil Pumps (3-50% pumps)	3	456 GPM/11 HP each	CS		

Table 1-4 Equipment Cost Estimates

	Selexol	SIMTECHE	
		2-Stage Promoter	*1-Stage Promoter*
CO_2 Separation Ratio	90%	90%	90%
Total Installed Cost, $ 2Q 2005			
Vessels, Exchangers, Pumps	$ 39,987,946	$ 48,049,292	$ 65,113,854
Refrigeration Compression	$ 3,046,633	$ 19,773,269	$ 28,098,849
CO_2 + Other Compression	$ 17,817,452	$ 8,323,126	$ 6,291,840
MDEA AGR	Not Required	$ 555,000	$ 555,000
Column Packing, Initial Fill	$ 1,069,338	$ 67,874	$ 71,762
Total CO2 Removal and Compression	**$61,921,368**	**$76,768,561**	**$100,131,304**

Table 1-5 Power Production Summary

Power Production Summary	Case		
	Selexol	SIMTECHE	
		2-Stage Promoter	*1-Stage Promoter*
CO_2 Separation Ratio	90%	90%	90%
Gross Plant Power, kW_e	474,689	497,392	497,392
Auxilliary Power Loads			
CO_2 Capture and Compression	35,766	44,265	59,928
Power Plant	54,842	54,878	54,878
Net Plant Power, kW_e	384,081	398,249	382,586
Net Efficiency, %HHV	35.9	36.5	35.1
Heat Rate, BTU/kWhr	9,691	9,347	9,729
Capture Parasitic Load, %	7.5%	8.9%	12.0%

Table 1-6 Carbon Control Cost

Cases	No Capture	Selexol	SIMTECHE	
			2-Stage Promoter	*1-Stage Promoter*
CO_2 Separation Ratio	0%	90%	90%	90%
Onstream Factor	80%	80%	80%	80%
Coal Feed (as received), Short Tons per Year	1,009,152	1,118,068	1,118,068	1,118,068
Carbon Dioxide Captured, STPY		2,288,719	2,278,145	2,278,499
Carbon Dioxide Emitted, STPY	2,353,363	237,382	247,995	252,624
Cost of Production, $ per Year				
Capital Charge	$ 87,113,705	$ 112,484,409	$ 115,290,529	$ 119,706,087
Coal Cost	$ 29,265,408	$ 32,423,972	$ 32,423,972	$ 32,423,972
Chemicals/Consumables	$ 2,271,894	$ 2,353,786	$ 2,285,469	$ 2,285,469
Maintenance Cost, Materials&Labor	$ 9,218,382	$ 11,903,112	$ 12,200,056	$ 12,667,311
Operating Labor and Administration	$ 8,688,493	$ 8,830,182	$ 8,830,182	$ 8,830,182
Total Cost of Production, $ per year	**$ 136,557,882**	**$ 167,995,461**	**$ 171,030,207**	**$ 175,913,021**
Net Power Produced, kW_e	424,540	384,081	398,249	382,586
Cost of Electricity, cents per kWhr	4.59	6.24	6.13	6.56
% Increase in COE	**0%**	**36%**	**34%**	**43%**
Avoided Capture Cost, $ per Ton of CO_2	**0**	**$ 25**	**$ 24**	**$ 31**
Avoided Capture Cost, $ per Ton of Carbon	0	**$ 6.89**	**$ 6.44**	**$ 8.34**
Capture Cost of Shift		$ 6.18	$ 5.91	$ 5.96

Task 3 - Supercritical CO2 Recovery

Objective

This task has the objective to determine if pumping the hydrate slurry to supercritical conditions of 2200 psia is a cost effective way of eliminating the CO_2 sequestration compression costs. The CO2 sequestration compressor is a major cost item and source of energy consumption for the Simteche CO_2 removal process. By taking the CO_2 hydrate slurry recovered from the hydrate formation reactors and pumping it to sequestration delivery pressures of 2200 psia, the CO_2 hydrate can be heated and the CO_2 separated at supercritical conditions and sent to sequestration without any additional compression. This eliminates compression of the combined dried CO_2 product gas as is done in the once-through 1000 psia base case.

Summary

There is no savings in pumping CO_2 hydrate to its sequestration pressure compared to compressing the produced CO2 to its sequestration pressure. This is because supercritical CO_2 separation must occur at 100 ^oF, and as a result, the refrigeration load of the recycle water is much higher than that of the base case.

Techno-economic analysis comparing supercritical CO_2 recovery with the current method was performed on a fully integrated IGCC (EPRI prototype) plant with the once-through Simteche's CO_2 hydrate capture process. This was done because there is not only a significant capital cost difference, but a significant operating cost difference in the two plants which affects the net revenue.

Table 1-1 summarizes the major differences, and the final economic comparison on a cost of electricity basis. As can be seen, the supercritical CO_2 recovery case ends up with a higher capital cost by over $25 million as well as a cost of electricity 0.33 cents/kWhr more than the base case. There is a significant increase in refrigeration compression loading associated with the supercritical CO_2 recovery case.

The analysis was performed on a once-through Simteche hydrate process at 68% CO_2 recovery. The outcome, however, is not expected to be different at a higher CO_2 (i.e., 90%) recovery.

Supercritical CO_2 Recovery

The supercritical flow scheme is shown in Figures 2-1 and 2-2. In this variation of the once-through Simteche CO_2 recovery process, the resultant liquid hydrate slurry from the hydrate formation reactor is pumped from 940 psia to CO_2 sequestration pressures of 2200 psia. The hydrate is first heated to 54^oF to disassociate the CO_2 and H_2S from their hydrates and then further heated to 100^oF to facilitate gravity separation of the CO_2 from the recycle water. The first heating is accomplished by condensing part of the 1st stage NH3 refrigeration compressor effluent. The second heating is accomplished by condensing the second stage NH3 refrigeration compressor effluent. The hydrate needs to be heated to at least 100 ^oF to achieve a reasonable gravity separation of the dense phase supercritical CO_2 and water due to the small density differences between CO_2 and water at temperatures below 100^oF. In addition, because water is the dominant continuous phase, the CO_2 must rise through the more viscous water phase, making the separation easier as the temperature increases and the continuous phase becomes less viscous.

The separated water containing small amounts of CO_2 and H_2S is letdown in pressure through a hydraulic recovery turbine to 750 psig. Here a flash gas containing CO_2 and H_2S is flashed off and mixed with acid gas from the Acid Gas Recovery Unit and compressed to CO_2 sequestration pressures of 2200 psia. A hydraulic recovery turbine is used to recover some of the high pressure energy of the recycle water and augment the motor drive of the Hydrate pump. Liquid from the 750 psig flash is then exchanged with hydrate formation reactor vapor to cool it. The water is then chilled to 34 ^oF and pumped back to the inlet of the Hydrate Formation Reactor.

The CO_2 separated from the heated hydrate at 2200 psia and 100 ^oF is mixed with compressed flash gas. The resultant dense phase material is sent to a Glycerol Dehydration unit to remove any residual water before being sent to CO_2 sequestration. Glycerol dehydration has been used successfully in some of the supercritical CO_2 pipeline systems since there is minimal solvent loss in the CO_2 compared to other glycols (Note 3).

Methodology

In order to evaluate this option, a simulation model was required. The current spreadsheet that was used for the previous tasks is incapable of handling the hydrate heating at supercritical conditions. As a result, a simulation model for the Supercritical CO_2 recovery plant was built using Hysys to develop the heat and mass balance and facilitate sizing and costing the plant. Hysys was modified to handle the flashes below critical pressure using Henry's law constants that were employed in the previous cases presented in the Phase II Engineering Analysis Report. The total CO_2 heats of hydrate formation (vapor phase to solid hydrate) was assumed to be 62 kJ/gmol. This is based upon a consensus of experimental data (Phase 1 report p. 5-14)(See appendix G Tables A-1 & A-2). Heat of hydrate formation of CO_2 and H_2S (solid hydrate to vapor phase) were assumed to be the same at supercritical pressures of 2200 psia as at 1000 psia. The H_2S heats of hydrate formation (solid hydrate to liquid phase) were assumed to be the same as CO_2 on a molar basis since they had similar hydration numbers and any deviation was expected to be small since the major contributor is the heat of fusion of water. In addition the heat of hydration of H2S represents only 2% of the hydrate heater duty. These heats of hydrate formation will need to be verified by experimental data.

The Hysys enthalpies for CO_2 in the supercritical range were in close agreement with published experimental data (Note1). The Hysys CO_2 densities in the supercritical region also agreed closely with published data in this region. This needs to be further investigated for the mixtures of CO_2 with H2S and other dissolved components.

After reviewing the densities of liquid CO_2 and water in the supercritical region, it was determined that the CO_2 water mixture must be heated to a least 100 ^oF to get a reasonable density difference to effect a good gravity separation between the CO_2 phase and the water phase. At these conditions the density of CO_2 changes dramatically. In addition, because water is the continuous phase its higher viscosity makes it difficult for the CO_2 phase to separate from the water. For example if the separation took at place at 60 ^oF, it would require 4 separators 30 inches in diameter 1400 ft long. At 80 ^oF, it would require 4 separators 3 ft in diameter 415ft long. At 100 ^oF requires 4 separators 6 ft in diameter 36ft long.

The solubility of CO_2 and H_2S in the supercritical water phase needs to be verified by experimental data.

The Supercritical CO_2 recovery cases, along with its comparison with the other cases, are presented in the following:

Table 1-1 - Economic Comparative Analysis
Table 1-2 - Equipment Cost Estimates
Table 1-3 - Power Production Summary
Table 1-4 - Carbon Control Cost
Table 1.5 - Utility summary
Table 1.6 - Equipment list
Table 1.7- Material Balance
Figure 2.1 - Simplified PFD CO_2 Recovery Section
Figure 2.2 - Simplified PFD NH_3 Refrigeration Section

Results

One of the first things we discovered in investigating this case was that the hydrate slurry had to be heated to at least 48 ^oF (Note 2) to release CO_2 and H_2S from their hydrates at 2200 psia. At these conditions, the density differences between CO_2 and water were too small to effectively separate these fluids by gravity. In addition the high viscosity of water at 48 ^oF added to the difficulty of gravity separation. It was necessary to heat the hydrate slurry to 100^oF to cost effectively separate the two phases by gravity. As a result there was a large increase in the refrigeration load to chill the recovered water to be recycled back to the Hydrate Formation reactor at 34^oF.

Another observation was that the energy required to pump the hydrate slurry to supercritical conditions was almost as much as that required for the CO_2 sequestration compressor in the once-through 1000 psia base case. This is because the hydrate slurry quantity is almost six times the mass of the product CO_2 that is compressed in the base case. In addition, the final stage of the CO_2 sequestration compressor is practically a pump since it operates in the supercritical region where the CO_2 density is similar to a liquid.

A summary of the net power production and the cost of electricity from the IGCC complex for the Once-thru Supercritical CO_2 case and the base case are compared in Table 1-1. This shows that there is a 10 MW power generation loss with the Supercritical CO_2 case compared to the once-thru base case. In addition, the cost of electricity increases to 6.08 cents from 5.75/kWhr. This is mainly due to the increase refrigeration load required to chill the recycle water from 100 ^oF.

Table 1-2 shows that the capital cost of the supercritical case is almost 50% higher than the base case. It has higher compressor costs as well as higher vessel and exchanger costs. Table 1-3 shows that the supercritical case net efficiency is lower than the once - thru base case 36.8% to 37.7% and the parasitic load is 8% compared to 6% for the base case.

Table 1-4 shows that the avoided capture cost of the supercritical case is $33/Ton of CO_2 compared to $25/Ton for the once - thru base case. The percent increase cost of electricity from no capture is 33% for the supercritical case and 25% for the base case.

Even if the hydrate could be separated at a temperature less than 100 ^oF, this option wouldn't be cost effective. There is almost no energy savings in pumping the hydrate to sequestration pressures compared to compressing the flash gases to the same pressure because the mass of the hydrate is six times the mass of the flashed gases.

Notes:

1. S. Angus, B. Armstrong, K.M. de Reuck, V.V. Altunin, O.G. Gadetskii, G.A. Chapela and J.S. Rowlinson, International Thermodynamic Tables of the Fluid State-Carbon Dioxide, IUPAC, Pergamon Press, Oxford 1976

2. K.Y.Soong, Riki Kobayashi, Water Content of CO_2 in Equilibrium with Liquid Water and/or Hydrates. SPE Dec. 1987

3. C.B. Wallace, Dehydration of Supercritical CO_2, 1985 Laurence Reid Gas Conditioning Conference**.**

Table 1-1 Comparison

Case		Once-Thru Case	Supercritical CO2
CO2 Recovery		68%	68%
Capital Cost (Note 1)	MM$	51.63	76.73
Power Revenues			
Gas Turbine Power		342295	342295
Steam Turbine Power		150314	150314
Generator Loss		(7265)	(7265)
Turboset Power		485344	485344
Fuel Gas Expander Power		10835	10835
Total Gross Power 'Generated	KW	496179	496179
Auxiliary Loads			
1st Stg NH3 Compressors	KW	16819	21012
2nd Stg NH3 Compressors	KW	3139	13921
Acid Gas Compressors	KW	152	157
Flash Gas Compressors	KW	1040	398
Hydrate Pump	KW		4766
Hydrate Hydraulic Recov Turb	KW		(2374)
CO2 Sequestration Compressor	KW	5289	
Cooling Water Pumps	KW	313	551
MDEA AGR	KW	55	
Recycle Water Pumps	KW	2662	1058
Dryer Regen Heater	KW	170	
Other Loads		9	144
CO2 Removal Plant Loads	KW	29648	39633
Cooling Tower Fans	KW	1141	1176
Balance of Plant	KW	53680	53680
Total Auxiliary loads	KW	84469	94489
OnStream Factor		0.8	0.8
Net Power	KW	411710	401690
Cost of Electricity	Cents/ kWhr	5.75	6.08

Table 1-2 Equipment Cost Estimates

CO_2 Removal and Compression	Selexol	SIMTECHE	
		Once-Thru Supercritical	*Once-Thru*
CO_2 Separation Ratio	90%	68%	68%
Total Installed Cost, $ 2Q 2005			
Vessels, Exchangers, Pumps	$ 39,987,946	$ 55,383,951	$ 32,227,217
Refrigeration Compression	$ 3,046,633	$ 17,762,975	$ 12,999,977
CO_2 + Other Compression	$ 17,817,452	$ 2,970,459	$ 5,795,199
MDEA AGR	Not Required	$ 584,000	$ 584,000
Column Packing, Initial Fill	$ 1,069,338	$ 30,000	$ 24,521
Total CO2 Removal and Compression	**$61,921,368**	**$76,731,385**	**$51,630,914**

Table 1-3 Power Production Summary

Power Production Summary	No Capture	Selexol	SIMTECHE	
			Once-Thru Supercritical	*Once-Thru*
CO_2 Separation Ratio	0%	90%	68%	68%
Gross Plant Power, kW_e	474,040	480,319	496,179	496,179
Auxilliary Power Loads				
CO_2 Capture and Compression		35,766	39,633	29,648
Power Plant	49,500	54,894	54,856	54,821
Net Plant Power, kW_e	424,540	389,659	401,690	411,710
Net Efficiency, %HHV	43.0	35.7	36.8	37.7
Heat Rate, BTU/kWhr	7,936	9,553	9,266.6	9,041
Capture Parasitic Load, %		7.4%	8.0%	6.0%

Table 1-4 Carbon Control Cost

Carbon Control Costs	No Capture	Selexol	SIMTECHE Once-Thru Supercritical	SIMTECHE Once-Thru
CO_2 Separation Ratio	0%	90%	68%	68%
Onstream Factor	80%	80%	80%	80%
Coal Feed (as received), Short Tons per Year	1,009,152	1,118,068	1,118,068	1,118,068
Carbon Dioxide Captured, STPY		2,288,719	1,724,240	1,724,192
Carbon Dioxide Emitted, STPY	2,353,363	237,382	801,947	801,947
Cost of Production, $ per Year				
Capital Charge	$ 87,113,705	$ 112,484,409	$ 115,283,502	$ 110,539,513
Coal Cost	$ 29,265,408	$ 32,423,972	$ 32,423,972	$ 32,423,972
Chemicals/Consumables	$ 2,271,894	$ 2,353,786	$ 2,491,894	$ 2,276,798
Maintenance Cost, Materials&Labor	$ 9,218,382	$ 11,903,112	$ 12,199,312	$ 11,697,303
Operating Labor and Administration	$ 8,688,493	$ 8,830,182	$ 8,830,182	$ 8,830,182
Total Cost of Production, $ per year	**$136,557,882**	**$ 167,995,461**	**$171,228,863**	**$ 165,767,768**
Net Power Produced, kW_e	424,540	389,659	401,690	411,710
Cost of Electricity, cents per kWhr	4.59	6.15	6.08	5.75
%Increase in Cost of Electricity		34%	33%	25%
Avoided Capture Cost, $ per Ton of CO2	**$ -**	**$ 24**	**$ 33**	**$ 25**
Avoided Capture Cost, $ per Ton of Carbon	**$ -**	**$ 6.5**	**$ 8.9**	**$ 6.8**
Capture Cost of Shift,$ per Ton of Carbon		$ 6.16	$ 7.84	$ 7.80

Table 1-5 Utilities Summary

Item No	Item Name	Load BHP		Elect. Power	Cooling Water		Refrigeration		Fuel MMBTU / HR	
		Norm.	Max (3).	KW	CW, MMbtu/hr	C.W. circ. GPM (2)	NH3, MMbtu/hr	Tons of Refrig.	Heat Abs.	Total Fired Duty
	Exchangers									
	Syngas Chiller						24.90	2,075		
	Hydrate Flash Reactor A									
	Hydrate Flash Reactor B									
	Recycle Water Chiller						159.0	13,250		
	Flash Gas Cooler				4.78	478				
						-				
	Acid Gas Compr 1st Intercooler				0.37	37				
	Acid Gas Compr 2nd Intercooler				0.17	17				
	Acid Gas Compr 3rd Intercooler				0.15	15				
	Ammonia Condenser				217.60	29,013				
	Lean Glycerol Cooler				1.72	172				
	Glycerol Regenerator								2.79	3.49
	Hydrate Formation Reactors									
	Hydrate Formation Reactor						328.82	27,402		
	Compressors									
	Flash Gas Compressor	507		398	0.03	3				
	Acid Gas Compressor	200		157	0.01	1				
	NH_3 Compressor - Stage 1	26,758		21,012	1.36	182				
	NH_3 Compressor - Stage 2	17,728		13,921	0.90	120				
	Pumps									
	Chilled Water Recycle Pumps	1,276		1,058						
	Cooling Water Pumps	701		551						
	Lube Oil Pumps	20		17						
	Demineralized Water Pumps	12.5		10.4						
	Hydrate Pump	6,069		4,766						
	Hydraulic Turbine	(3,350)		(2,374)						
	Lean Glycerol Pump	142		117						
	TOTAL	50,064		39,633	227.10	30,040	512.72	42,727	2.79	3.49

NOTES:
1 All Figures shown above represent normal utility usage requirements except:
() indicates normal utility make
* indicates intermittent usage or make, not included in totals
2 Cooling water supply temperature is 80 F. Makeup water to cooling tower is not shown
3 Utility consumption for max. load conditions is not shown.

Table 1-6 Equipment List

Equipment	No. ea.	Size/Duty	Material Tube/Shell	Temperature (°F) Tube/Shell	Pressure (psig) Tube/Shell
Reactors					
Hydrate Formation Reactors	10	329 MM Btu/hr total	304 SS/CS	200/200	1050/235
Hydrate Flash Reactor A	4	261 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/680
Hydrate Flash Reactor B	2	140 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/315
Exchangers					
Syngas Chiller	1	24.9 MM Btu/hr	304 SS/CS	200/200	1,100/235
Recycle Water Chiller	2	73.6. MM Btu/hr	304 SS/CS	200/200	300/235
Flash Gas Cooler	1	4.8 MM Btu/hr	CS/CS	285/200	680/50
Product Gas/Recycle Water Cooler.	1	10.8 MM Btu/hr	CS /304 SS	250/250	2310/800
Ammonia Condenser	4	217.6 MM Btu/hr	CS/CS	200/360	500/235
Acid Gas Compressor 1st Stage Intercooler	1	0.11 MM Btu/hr	CS/304 SS (Clad)	200/350	50/50
Acid Gas Compressor 2nd Stage Intercooler	1	0.05 MM Btu/hr	CS/304 SS (Clad)	200/350	50/110
Acid Gas Compressor 3rd Stage Intercooler	1	0.05 MM Btu/hr	CS/304 SS (Clad)	200/350	50/320
Rich/Lean Glycerol Exchanger	1	1.6 MM Btu/hr	CS/304 SS (Clad)	250/250	100/50
Lean Glycerol Cooler	1	1.7 MM Btu/hr	CS/CS	250/250	75/100
Glycerol Regenerator Condenser	1	0.2 MM Btu/hr	CS/CS	250/400	75/100
Vessels					
Syngas KO Drum	1	8'ID x 10' T-T	304 SS Clad	250	1,055
Slurry/Gas Separator	1	11.0'ID x 33.5' T-T	304 SS Clad	250	975
Flash Gas Separator	1	2.5'ID x 7.5' T-T	304 SS Clad	120	825
Acid Gas Compr KO Drum	1	1'ID x 5' T-T	304 SS Clad	160	50
Acid Gas 1st Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	100
Acid Gas 2nd Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	230
Acid Gas Product KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	670
Hydrate Separator	4	6.5'ID x 36.5' T-T	304 SS Clad	250	2310
Ammonia Separator	1	7.0'ID x 21' T-T	CS	200	235
Ammonia Surge Drum	1	11.5'ID x 40' T-T	CS	200	235
Ammonia Compr KO Drum	1	14.5'ID x40' T-T	CS	200	235
Glycerol Flash Drum	1	2.5'ID x7' T-T	CS	200	235
Columns					
Glycerol Contactor	1	5.5'ID x35' T-T 8 Trays	304 SS Clad CS 410 SS Trays	250	2300
Glycol Regenerator	1	2'ID x 157' T-T 4 SieveTrays	CS 410 SS Trays	500	50
Compressors					
Flash Gas Compressor	1	2,000 HP	CS	350	625
Acid Gas Compressor	1	75 HP	CS	350	610

NH3 Refrig 1st Stage Compressor	1	29,500 HP	CS	200	136
NH3 Refrig 2nd Stage Compressor	1	19,500 HP	CS	350	245
Equipment	No. ea.	Size/Duty	Material	Temperature (°F)	Pressure (psig)
			Tube/Shell	Tube/Shell	Tube/Shell
Pumps					
Chilled Water Recycle Pumps (3-50% pumps)	3	2725 GPM/ 710 HP each	304 SS	200	1100
Hydrate Pump (2-100%)	2	6,200 GPM/ 7000 HP each	304 SS	250	2400
Cooling Water Pumps (3-50% pumps)	3	12,000 GPM/350 HP each	CS	200	100
Demineralized Water Pumps (2-100% pumps)	2	0.14 GPM/0.03 HP each	CS		
Lube Oil Pumps (3-50% pumps)	3	975 GPM/24 HP each	CS		

Table 1-7 Material Balance

Coal Illinois No. 6

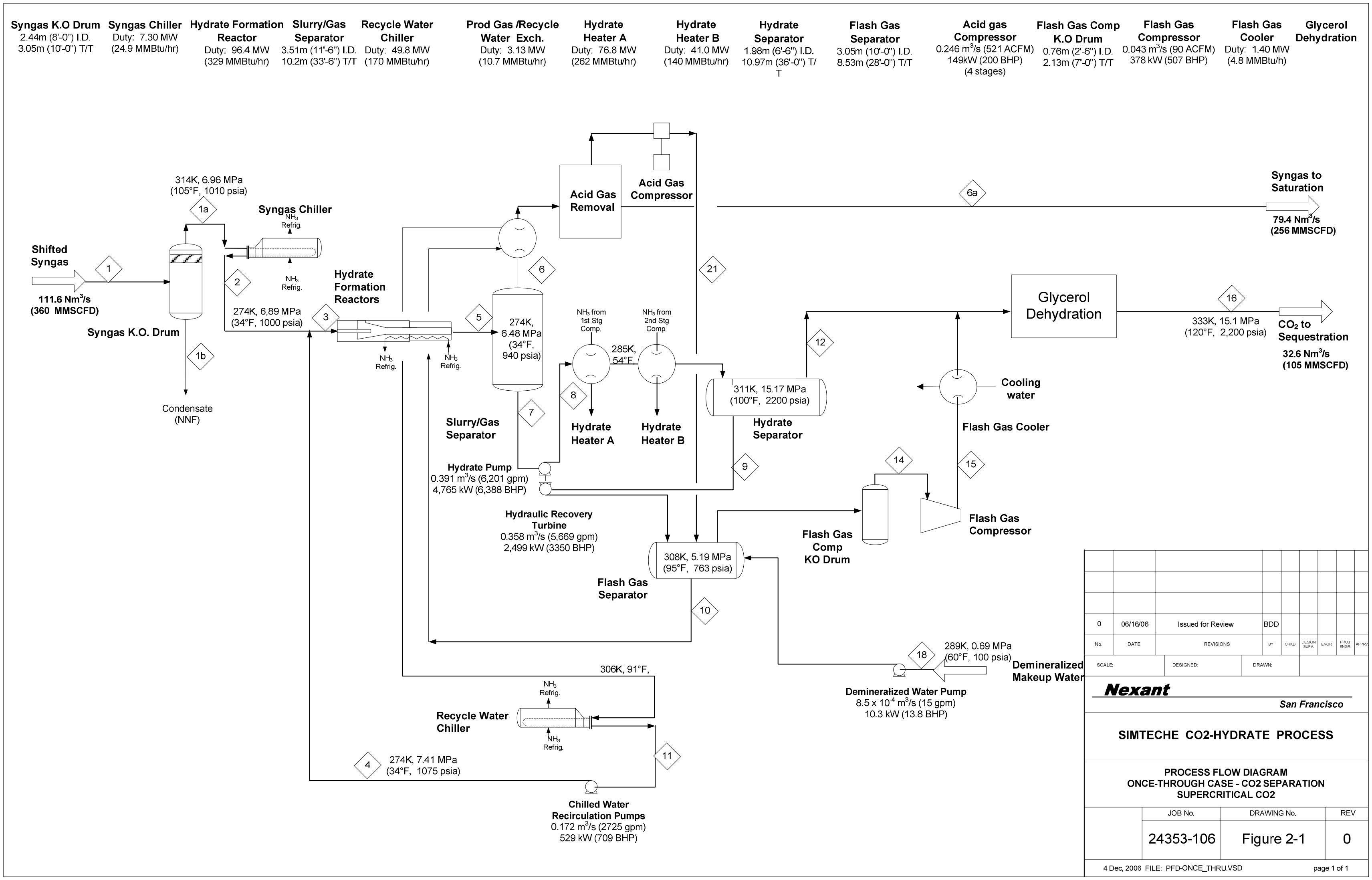
Wt% Sulfur 2.82 wt% dry (2.51 wt% S As Received

Stream		1		2		3		4		5		6	
Name		Sat'd Syngas to KO Drum		Chilled Syngas		Hydration Formation Reactor Feed		Recycle Water		Reactor Effluent		Treated Syngas	
Vapour Fraction		1		1		0.203		0				1	
Temperature (F)		105		34		35.74		35.28		34		34	
Pressure (psia)		1010		1000		1000		1075		940		940	
Molar Flow (lbmole/hr)		39558		39558		192703		153146		192703		28133	
Mass Flow (lb/hr)		811836		811836		3640043		2828207		3640043		314059	
Molecular Weight		20.5		20.5		18.9		18.5		18.9		11.2	
Z Factor		0.9524		0.9153				0.0580				0.9924	
Mass Density (lb/ft3)		3.6		4.2		15.5		64.5				2.0	
Actual Liquid Flow (USGPM)						5502		5469					
Viscosity (cP)		0.015		0.014				1.400				0.010	
Surface Tension (dyne/cm)						74.56		74.8		75.79			
Component	Mol.Wt.	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr
Hydrogen	2.02	21418.7	43176	21418.7	43176	21418.8	43176	0.0	0	21418.8	43176	21357.3	43052
CO2	44.01	16381.1	720930	16381.1	720930	18936.9	833407	2555.7	112477	6420.5	282566	5251.0	231095
H2S	34.08	248.4	8466	248.4	8466	437.3	14900	188.8	6434	39.3	1339	25.0	852
Argon	39.95	292.7	11694	292.7	11694	292.7	11694	0.0	0	292.7	11694	291.8	11658
Nitrogen	28.01	289.3	8105	289.3	8105	289.3	8105	0.0	0	289.3	8105	288.4	8080
CO	28.01	382.3	10709	382.3	10709	382.3	10709	0.0	0	382.3	10709	380.5	10659
Methane	16.04	537.8	8628	537.8	8628	537.8	8628	0.0	0	537.8	8628	533.9	8566
COS	60.07	0.04	2	0.04	2	0.0	2	0.0	0	0.0	2	0.0	1
Ammonia	17.03	3.4	58	3.4	58	218.4	3720	215.0	3662	218.4	3720	2.6	44
H2O	18.02	3.8	68	3.8	68	150190.0	2705703	150186.2	2705634	72704.2	1309780	2.8	50
CO2 Hydrate	152.10									12516.3	1903748		
H2S Hydrate	142.17									397.96	56576		
Total		39558	811836	39558	811836	192703	3640043	153146	2828207	192703	3640043	28133	314059

Stream		6A		6B		7		8		8A		8B		9	
Name		Treated Syngas to AGR		Treated Syngas from AGR		Slurry to Hydrate Pump		Slurry from Hydrate Pump		Slurry from Hydrate Heater A		Slurry to Hydrate Separator		Water from Hydrate Separator	
Vapour Fraction		1		1		0		0		0		0		0	
Temperature (F)		85		105		34		39.65		54		100		100	
Pressure (psia)		935		922		940		2225		2218		2210		2210	
Molar Flow (lbmole/hr)		28133		28056		164570		164570		164570		164570		153693	
Mass Flow (lb/hr)		314059		310951		3325984		3325984		3325984		3325984		2862010	
Molecular Weight		11.2		11.1		20.2		20.2		20.2		20.21		18.621	
Z Factor		1.0010		1.0030										0.1087	
Mass Density (lb/ft3)		1.8		1.7		66.8		66.4		64.0		61.1		63.0	
Actual Liquid Flow (USGPM)						6207		6245		6481		6789		5660	
Viscosity (cP)		0.011		0.011		1.555		1.448		1.243		0.803		0.677	
Surface Tension (dyne/cm)														68.16	
Component	Mol.Wt.	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr
Hydrogen	2.02	21357	43052	21357.3	43052	61.4	124	61.4	124	61.4	124	61.4	124	1.67	3
CO2	44.01	5251	231095	5200.1	228853	1169.5	51470	1169.5	51470	13685.9	602312	13685.9	602312	3477.636	153050
H2S	34.08	25	852	1.1	37	14.3	487	14.3	487	412.2	14048	412.2	14048	188.7469	6432
Argon	39.95	292	11658	291.8	11658	0.9	36	0.9	36	0.9	36	0.9	36	0.0006	0
Nitrogen	28.01	288	8080	288.4	8080	0.9	25	0.9	25	0.9	25	0.9	25	0.0648	2
CO	28.01	381	10659	380.5	10659	1.8	50	1.8	50	1.8	50	1.8	50	0.0279	1
Methane	16.04	534	8566	533.9	8566	3.9	62	3.9	62	3.9	62	3.9	62	0.0003	0
COS	60.07	0	1	0.0	1	0.0	1	0.0	1	0.0	1	0.0	1	0.0003	0
Ammonia	17.03	3	44	2.6	44	215.8	3675	215.8	3675	215.8	3675	215.8	3675	214.4004	3651
H2O	18.02	3	50	0.0	0	72701.4	1309730	72701.4	1309730	150187.2	2705652	150187.2	2705652	149810.8	2698871
CO2 Hydrate	152.10					12516.3	1903748	12516.3	1903748		0		0		0
H2S Hydrate	142.17					398.0	56576	398.0	56576		0		0		0
Total		28133	314059	28056	310951	164570	3325984	164570	3325984	164570	3325984	164570	3325984	153693	2862010

Stream Number		9A		10		10A		11		12		14	
Name		Water from Hydraulic Turbine		Flash Separator Liquid		Recycle Water to Chiller		Recycle Water from Chiller		CO2 from Hydrate Separator		Flash Gas	
Vapour Fraction		0.0062		0		0		0		0		1	
Temperature (F)		94.9		95.0		91.2		34.0		100.0		95.0	
Pressure (psia)		763		763		758		748		2210		763	
Molar Flow (lbmole/hr)		153693		153146		153146		153146		10877		1001	
Mass Flow (lb/hr)		2862010		2828207		2828207		2828207		463974		43685	
Molecular Weight		18.6		18.5		18.5		18.5		42.7		43.6	
Z Factor				0.0377		0.0376		0.0405		0.3065		0.6880	
Mass Density (lb/ft3)		57.3		62.8		62.9		64.5		51.2		8.1	
Actual Liquid Flow (USGPM)				5611		5602		5470		1129			
Viscosity (cP)				0.72		0.75		1.42		0.06		0.02	
Surface Tension (dyne/cm)		69.06		69.06		69.41		74.93		2.641			
Component	Mol.Wt.	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr
Hydrogen	2.02	1.7	3	0.0	0	0.0	0	0.0	0	59.8	120	1.6	3
CO2	44.01	3477.6	153050	2555.7	112476	2555.7	112476	2555.7	112476	10208.2	449262	972.9	42816
H2S	34.08	188.7	6432	188.8	6434	188.8	6434	188.8	6434	223.5	7616	23.9	813
Argon	39.95	0.0	0	0.0	0	0.0	0	0.0	0	0.9	36	0.0	0
Nitrogen	28.01	0.1	2	0.0	0	0.0	0	0.0	0	0.8	23	0.1	2
CO	28.01	0.0	1	0.0	0	0.0	0	0.0	0	1.7	49	0.0	1
Methane	16.04	0.0	0	0.0	0	0.0	0	0.0	0	3.9	62	0.0	0
COS	60.07	0.0	0	0.0	0	0.0	0	0.0	0	0.0	1	0.0	0
Ammonia	17.03	214.4	3651	214.3	3650	214.3	3650	214.3	3650	1.4	24	0.1	1
H2O	18.02	149810.8	2698871	150186.9	2705646	150186.9	2705646	150186.9	2705646	376.4	6781	2.7	49
CO2 Hydrate	152.10												
H2S Hydrate	142.17												
Total		153693.3	2862010	153145.8	2828207	153145.8	2828207	153145.8	2828207	10876.6	463973.7	1001.3	43685

Stream Number		15		15A		16		18		18A		21	
Name		Flash Gas Compressor outlet		Flash Gas Cooler outlet		Feed to Glycerol Dehydration		Demin Water Feed		Demin water pump outler		Acid Gas from AGR Unit	
Vapour Fraction		1		0		0		0		0		1	
Temperature (F)		274.2		100.0		100.0		60.0		62.6		267	
Pressure (psia)		2215		2210		2210		100		765		765	
Molar Flow (lbmole/hr)		1001		1001		11878		376		376		78	
Mass Flow (lb/hr)		43685		43685		507659		6774		6774		3108	
Molecular Weight		43.6		43.6		42.7		18.0		18.0		40.0	
Z Factor		0.7651		0.3336		0.3085		0.0051		0.0388		0.8746	
Mass Density (lb/ft3)		16.0		48.1		51.0		63.3		63.354		4.4882	
Actual Liquid Flow (USGPM)				113		1242		13.3		13.3			
Viscosity (cP)		0.03		0.05		0.06		1.12		1.08		0.02	
Surface Tension (dyne/cm)				0.4467		2.456		73.73		73.48			
Component	Mol.Wt.	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr	lbmol/hr	lb/hr
Hydrogen	2.02	1.6	3	1.6	3	61.4	124	0	0	0	0	0	0
CO2	44.01	972.9	42816	972.9	42816	11181.1	492077	0	0	0	0	50.9452	2242
H2S	34.08	23.9	813	23.9	813	247.4	8429	0	0	0	0	23.9319	815
Argon	39.95	0.0	0	0.0	0	0.9	36	0	0	0	0	0	0
Nitrogen	28.01	0.1	2	0.1	2	0.9	25	0	0	0	0	0	0
CO	28.01	0.0	1	0.0	1	1.8	50	0	0	0	0	0	0
Methane	16.04	0.0	0	0.0	0	3.9	62	0	0	0	0	0	0
COS	60.07	0.0	0	0.0	0	0.0	1	0	0	0	0	0	0
Ammonia	17.03	0.1	1	0.1	1	1.5	25	0	0	0	0	0	0
H2O	18.02	2.7	49	2.7	49	379.1	6830	376	6774	376	6774	2.7935	50
CO2 Hydrate	152.10												
H2S Hydrate	142.17												
Total		1001.3	43685	1001.3	43685	11877.9	507659	376	6774	376	6774	77.6706	3108


Syngas K.O Drum
2.44m (8'-0") I.D.
3.05m (10'-0") T/T
Syngas Chiller
Duty: 7.30 MW
(24.9 MMBtu/hr)
Hydrate Formation Reactor
Duty: 96.4 MW
(329 MMBtu/hr)
Slurry/Gas Separator
3.51m (11'-6") I.D.
10.2m (33'-6") T/T
Recycle Water Chiller
Duty: 49.8 MW
(170 MMBtu/hr)
Prod Gas /Recycle Water Exch.
Duty: 3.13 MW
(10.7 MMBtu/hr)
Hydrate Heater A
Duty: 76.8 MW
(262 MMBtu/hr)
Hydrate Heater B
Duty: 41.0 MW
(140 MMBtu/hr)
Hydrate Separator
1.98m (6'-6") I.D.
10.97m (36'-0") T/T
Flash Gas Separator
3.05m (10'-0") I.D.
8.53m (28'-0") T/T
Acid gas Compressor
0.246 m3/s (521 ACFM)
149kW (200 BHP)
(4 stages)
Flash Gas Comp K.O Drum
0.76m (2'-6") I.D.
2.13m (7'-0") T/T
Flash Gas Compressor
0.043 m3/s (90 ACFM)
378 kW (507 BHP)
Flash Gas Cooler
Duty: 1.40 MW
(4.8 MMBtu/h)
Glycerol Dehydration
Shifted Syngas
111.6 Nm3/s
(360 MMSCFD)
Syngas K.O. Drum
314K, 6.96 MPa
(105°F, 1010 psia)
Syngas Chiller
NH3 Refrig.
274K, 6.89 MPa
(34°F, 1000 psia)
Hydrate Formation Reactors
Condensate
(NNF)
Acid Gas Removal
Acid Gas Compressor
Syngas to Saturation
79.4 Nm3/s
(256 MMSCFD)
274K, 6.48 MPa
(34°F, 940 psia)
NH3 from 1st Stg Comp.
NH3 from 2nd Stg Comp.
285K, 54°F
311K, 15.17 MPa
(100°F, 2200 psia)
Glycerol Dehydration
333K, 15.1 MPa
(120°F, 2,200 psia)
CO2 to Sequestration
32.6 Nm3/s
(105 MMSCFD)
Cooling water
Flash Gas Cooler
Slurry/Gas Separator
Hydrate Heater A
Hydrate Heater B
Hydrate Separator
Hydrate Pump
0.391 m3/s (6,201 gpm)
4,765 kW (6,388 BHP)
Hydraulic Recovery Turbine
0.358 m3/s (5,669 gpm)
2,499 kW (3350 BHP)
308K, 5.19 MPa
(95°F, 763 psia)
Flash Gas Separator
Flash Gas Comp KO Drum
Flash Gas Compressor
289K, 0.69 MPa
(60°F, 100 psia)
Demineralized Makeup Water
Demineralized Water Pump
8.5 x 10-4 m3/s (15 gpm)
10.3 kW (13.8 BHP)
306K, 91°F,
Recycle Water Chiller
274K, 7.41 MPa
(34°F, 1075 psia)
Chilled Water Recirculation Pumps
0.172 m3/s (2725 gpm)
529 kW (709 BHP)
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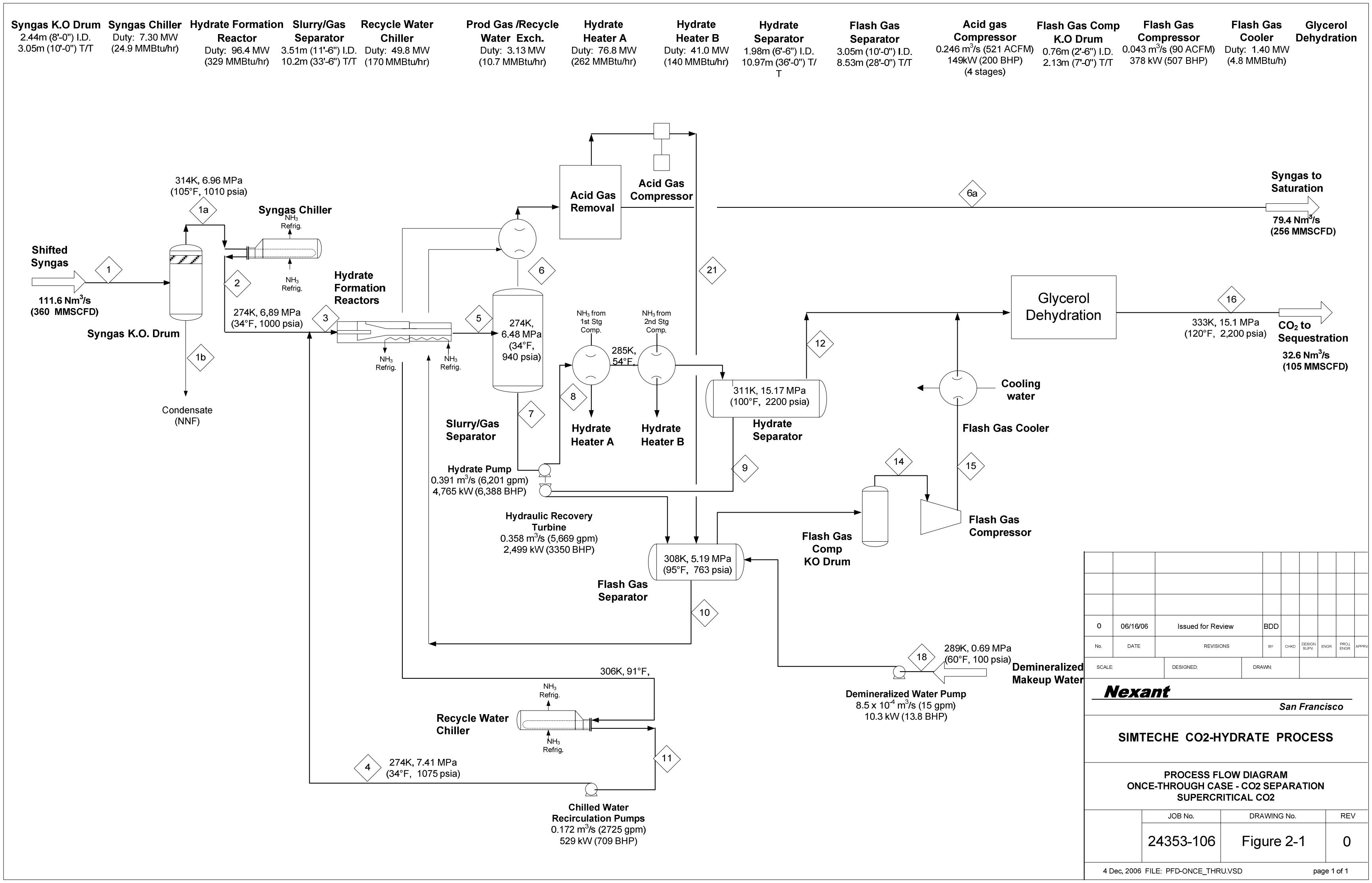
No. DATE REVISIONS BY CHKD DESIGN SUPV ENGR PROJ ENGR APPR
SCALE: DESIGNED: DRAWN:
Nexant
San Francisco
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
ONCE-THROUGH CASE - CO2 SEPARATION
SUPERCRITICAL CO2
JOB No. DRAWING No. REV
24353-106 Figure 2-1 0
4 Dec, 2006 FILE: PFD-ONCE_THRU.VSD page 1 of 1



Syngas K.O Drum
2.44m (8'-0") I.D.
3.05m (10'-0") T/T
Syngas Chiller
Duty: 7.30 MW
(24.9 MMBtu/hr)
Hydrate Formation Reactor
Duty: 96.4 MW
(329 MMBtu/hr)
Slurry/Gas Separator
3.51m (11'-6") I.D.
10.2m (33'-6") T/T
Recycle Water Chiller
Duty: 49.8 MW
(170 MMBtu/hr)
Prod Gas /Recycle Water Exch.
Duty: 3.13 MW
(10.7 MMBtu/hr)
Hydrate Heater A
Duty: 76.8 MW
(262 MMBtu/hr)
Hydrate Heater B
Duty: 41.0 MW
(140 MMBtu/hr)
Hydrate Separator
1.98m (6'-6") I.D.
10.97m (36'-0") T/T
Flash Gas Separator
3.05m (10'-0") I.D.
8.53m (28'-0") T/T
Acid gas Compressor
0.246 m3/s (521 ACFM)
149kW (200 BHP)
(4 stages)
Flash Gas Comp K.O Drum
0.76m (2'-6") I.D.
2.13m (7'-0") T/T
Flash Gas Compressor
0.043 m3/s (90 ACFM)
378 kW (507 BHP)
Flash Gas Cooler
Duty: 1.40 MW
(4.8 MMBtu/h)
Glycerol Dehydration
Shifted Syngas
111.6 Nm3/s
(360 MMSCFD)
Syngas K.O. Drum
314K, 6.96 MPa
(105°F, 1010 psia)
Syngas Chiller
Condensate
(NNF)
274K, 6,89 MPa
(34°F, 1000 psia)
Hydrate Formation Reactors
274K,
6.48 MPa
(34°F,
940 psia)
Acid Gas Removal
Acid Gas Compressor
Syngas to Saturation
79.4 Nm3/s
(256 MMSCFD)
Glycerol Dehydration
333K, 15.1 MPa
(120°F, 2,200 psia)
CO2 to Sequestration
32.6 Nm3/s
(105 MMSCFD)
285K,
54°F
311K, 15.17 MPa
(100°F, 2200 psia)
Hydrate Separator
Hydrate Heater A
Hydrate Heater B
Slurry/Gas Separator
Cooling water
Flash Gas Cooler
Hydrate Pump
0.391 m3/s (6,201 gpm)
4,765 kW (6,388 BHP)
Hydraulic Recovery Turbine
0.358 m3/s (5,669 gpm)
2,499 kW (3350 BHP)
308K, 5.19 MPa
(95°F, 763 psia)
Flash Gas Separator
Flash Gas Comp KO Drum
Flash Gas Compressor
289K, 0.69 MPa
(60°F, 100 psia)
Demineralized Makeup Water
Demineralized Water Pump
8.5 x 10-4 m3/s (15 gpm)
10.3 kW (13.8 BHP)
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SUPERCRITICAL CO2
JOB No. 24353-106
DRAWING No. Figure 2-1
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4 Dec, 2006 FILE: PFD-ONCE_THRU.VSD
page 1 of 1

Task 4 – 3000 Psia Hydrate Formation Reactor Operating Pressure

Objective

One project goal for the Simteche Process is to find a process scheme that can achieve 90% CO_2 recovery in a single stage. This is theoretically possible by increasing the CO_2 partial pressure in the syngas feed. Preliminary calculations have shown that this can be achieved at hydrate reactor operating conditions of 3,000 psia and 34 ^oF.

The objective of this task is revise the process scheme to allow the hydrate formation reactor to be operated at 3000 psia and 34^oF and to determine the capital and operating costs for the process. The results can then be compared to the 2 stage promoter case operation and also to the optimized Selexol process which are designed to recover 90% CO_2. It is also compared to the once thru case at 1000 psia and 34 ^oF which recovered 68%

Summary

There is a considerable increase in operating and capital costs associated with improving the CO_2 recovery from 68 to 90 % by operating the hydrate formation reactor at 3000 psia and 34^oF. This increase in costs makes the scheme economically less attractive (i.e., capital cost increases from $51.63 to $99.47 MM) -Table 1-1. The overall cost increase is mostly due to the following:

- Hydrate formation reactor has to operate at three times the pressure at which syngas is required to be delivered at the outlet of the plant.
- Higher operating pressure resulted in more expensive reactors, exchangers and vessels
- Additional equipment such as two stages of high pressure compression and expansion are required
- The recycle water pumping energy also became significant (i.e., 12.2 instead of 2.7 MW).
- CO_2 must be recovered at the same low pressures as the base case, so there are no savings in CO_2 compression costs from the base case

Techno-economic analysis of this 3000 psia Hydrate Formation reactor CO_2 recovery was performed on a fully integrated IGCC (EPRI prototype) plant to facilitate comparison with the Once-through Simteche's CO_2 hydrate capture process, as well as other case studies. This was done to ensure that the capital costs, utilities and other operating costs were compared on a consistent basis.

Table 1-1 summarizes the major energy differences between the cases as well as the cost of electricity. As can be seen, the 3000 psia Hydrate Formation Reactor case ends up with a higher capital cost and cost of electricity than all the other cases. Its capital cost is $99 million compared to $62 million for Selexol, $77 million for the 2 stage promoter case and $ 52 million for the once-thru base case at 1000 psia reactor. Its cost of electricity is 6.65 cents versus 6.15 cents for Selexol, 6.13 cents for the 2 stage promoter and 5.75 cents for the once-thru base case at 1000 psia reactor.

When one compares this case with an optimized Selexol case, which also recovers 90% CO_2, Table 1-1 shows that the 3000 psia Hydrate Formation Reactor case has a capital cost that is about $35 million higher and a cost of electricity that is 0.41 cents/kWhr higher than Selexol (6.65 to 6.24 cents/kWhr).

3000 Psia Hydrate Formation Reactor Case – Process Scheme

In this high pressure variation of the once-through Simteche CO_2 recovery process, the syngas is compressed and inter-cooled in a two stage compressor to 3100 psia This is followed with cooling by exchange with cooling water, syngas product and ammonia to 34^{o}F (see Fig 2.1). It is then mixed with recycled chilled water and sent to the hydrate formation reactors. Here the mixture is chilled by exchange with ammonia and the resultant CO_2 and H_2S hydrates are formed. The reactor effluent is sent to a slurry/gas separator which operates at 2900 psia and 34^{o}F and where the phases are separated. The gas stream is heated first by exchange with Syngas feed and then by exchange with the first stage Syngas compressor discharge to preheat it to 148 ^{o}F before being sent to the first stage turbo expander. This preheat is sufficient to avoid any ice or other solids formation in the expander and downstream equipment. The effluent from the first stage expander exhausts at 1600 psia and is then exchanged with the first stage syngas compressor discharge to preheat it to 137 ^{o}F before being sent to the second stage turbo expander. The second stage expander exhausts at 922 psia where it is sent to an acid gas removal unit which removes H_2S to meet the syngas product specifications of 40 ppm sulfur. The final product is then sent to a syngas humidifier before going to the GE 7H combined cycle power plant. A small waste gas stream from the acid gas removal unit containing H_2S and a small amount of CO_2 is compressed to 630 psia.

The liquid slurry from the slurry/gas separator is letdown in pressure to 625 psia and the slurry is heated to 54 ^{o}F in hydrate flash reactor A. Here the hydrate is disassociated to CO_2, H_2S and water and 59% of the CO_2 is flashed off and sent to the mol sieve gas dehydration unit. The heating is accomplished by condensing part of the 1st stage NH_3 refrigeration compressor effluent. The remaining liquid containing water, CO_2 and H_2S is letdown in pressure to 285 psia and sent to hydrate flash reactor B. Here the mixture is heated again and additional CO_2 and H_2S are flashed off. This heating is also accomplished by condensing part of the 1st stage NH3 refrigeration compressor effluent. This off gas is compressed in a flash gas compressor to 630 psia and mixed with the off gas from hydrate flash reactor A and the waste gas from the acid gas removal unit. This mixture is then sent to a mol sieve gas dehydration unit to meet the product CO_2 dryness specification. The dried CO_2 is then compressed in the CO_2 sequestration compressor and cooled and sent offsite for sequestration.

The liquid from the Hydrate Flash Reactor B containing mostly water and dissolved CO_2 and H_2S is then chilled to 34 ^oF and pumped back to the inlet of the Hydrate Formation reactor.

Techno-economic Analysis - Methodology

An existing spreadsheet model, utilized in designing the cases included in the Engineering Analysis Report of February 2006, was used for to develop the material balances. It incorporated the Henry's law coefficients to handle the flashes with the hydrate present. An additional simulation model was built using Hysys to develop the heat and mass balance and facilitate sizing and costing the plant. The Hysys model was modified to handle the equilibrium flashes using the same Henry's law constants that were employed in the spreadsheet. The simulation model was unable to recognize hydrate formation or hydrate disassociation. The heat balance around the Hydrate Formation reactor and the Hydrate Heater were handled separately using Hysys heat capacities and heats of formation of the hydrates. The total CO_2 heats of hydrate formation (vapor phase to solid hydrate) was assumed to be 62 kJ/gmol. This is based upon a consensus of experimental data (Phase 1 report p. 5-14) (See appendix G Tables A-1 & A-2). Heat of hydrate formation of CO_2 and H_2S (solid hydrate to vapor phase) were assumed to be the same at 3000 psia as at 1000 psia. The H_2S heats of hydrate formation were assumed to be the same as CO_2 on a molar basis since they had similar hydration numbers and any deviation was expected to be small since the major contributor is the heat of fusion of water. In addition the heat of hydration of H2S represents only 2% of the hydrate formation reactor duty. These heats of hydration would need to be verified by experimental data should this concept be analyzed further.

The property packages used in Hysys are known to be accurate in the 1000 to 3000 psia conditions for syngas compression and expansion.

Because the high pressure flash takes place at 2900 psia, well above the syngas product delivery pressure of 900 psia, additional energy can be recovered from this stream by means of a gas turbo expander to reduce the syngas pressure to 900 psia. However, expanders can result in low discharge temperatures where ice and hydrates can form. To avoid the potential for low temperatures the expander feed gas is preheated with exchange with the Syngas compressor discharge streams. This has the following advantages:

- Reduces the cooling water requirement for the syngas compressors
- Increases the power attainable from the gas turbines
- Eliminates hydrate formation and localized freezing in the downstream acid gas removal unit

The 3000 psia operating conditions case along with its comparison with the other cases are presented in the following:

Results

Compared with Selexol - Table 1-1 show that operating the hydrate formation reactor at 3000 psia is not as cost effective as using the optimized Selexol process for CO_2 recovery. The Selexol option which also recovers 90% CO_2 has a capital cost that is $35.27 million less and annual operating revenue that is $2.8 million greater than the present 3000 psia scheme. Also, as can be seen, the 3000 psia Simteche process requires considerably more compressors and compression energy than Selexol, 52 MW to 29 MW and a 7 MW larger pumping requirement. Thus, even though the Selexol option generates 13 MW less gross power than the 3000 psia case, it still has a lower cost of electricity 6.24 cents/kWhr to 6.65 cents/kWhr.

Compared with Two-Stage Promoter Case 90% Recovery - Table 1-1 shows that the 3000 psia case has a capital cost $22.7 million greater than the simteche two stage promoter case. It also generates 21.2 MW less net power than the two stage promoter case. This is due to the higher refrigeration costs and the added cost of syngas compression and operating at 3 times the operating pressure. Because of higher capital and lower net power generation the 3000 psia case has a higher cost of electricity of 6.65 cents/kWhr vs 6.15cents/kWhr.

Compared with Base Case 68% Recovery - In attempting to increase the CO_2 recovery from 68 to 90% it becomes ever more difficult to get the incremental CO_2 recovery. As can be seen in Table 1-1, to increase the CO_2 recovery from 68 to 90 % requires a doubling of the energy consumption for the CO_2 recovery plant from 29,648 KW to 57,446 KW. Thus a 33% increase in CO_2 recovery results in a 94% increase in power consumption. The refrigeration load increased slightly less than the increase in CO_2 removal, but there was an additional 21 MW required for syngas compression although a credit of 7.6 MW was gained by utilizing the turbo expanders to recover energy from the high pressure gas. However, there was an incremental 9.5 MW required for the recycle water pump. In this task the syngas is compressed up to three times the pressure that is required to deliver it to the fuel gas saturation section.

As can be seen in Tables 1-1 and 1-4, the cost of electricity increases from 5.75 to 6.65 cents/kWhr to increase the CO2 recovery from 68 to 90%

Table 1-1 Economic Comparative Analysis

Case			3000 Psia Hydrate Form Reactor	2 Stage Promoter	1000 Psia Hydrate Form Reactor
CO2 Recovery		**Selexol 90%**	**90%**	**90%**	**68%**
CO2 Recovery and CompressionCapital Cost	MM$	61.92	99.47	76.77	51.63
Power Revenues					
Gas Turbine Power	KW	333720	341577	345355	342295
Steam Turbine Power (Note 1)	KW	139294	145814	150362	150314
Generator Loss	KW	(7330)	(7250)	(7330)	(7265)
Turboset Power	KW	465684	480141	488387	485344
Fuel Gas Expander Power	KW	9005	9162	9005	10835
Total Gross Power 'Generated	KW	474689	489303	497392	496179
Auxiliary Loads	KW				
1st Stg NH3 Compressors	KW	3787	19672	16819	16819
2nd Stg NH3 Compressors	KW		2869	2950	3139
Promoter 1st Stg NH3 Compressors	KW			9817	
Promoter 2nd Stg NH3 Compressors	KW			738	
Acid Gas Compressors	KW		46	122	152
Flash Gas Compressors	KW	9726	1308	1040	1040
Promoter Flash Gas Compressors	KW			1026	
1st stg Syngas Compressor	KW		10716		
2nd stg Syngas Compressor	KW		10443		
CO2 Sequestration Compressor	KW	15379.6	7044	6969	5289
1st Stg Product Gas Expander			(4026)		
2nd Stg Product Gas Expander			(3536)		
Solvent pumps	KW	5043			
Cooling Water Pumps	KW	501	459	443	313
Recycle Water Pump	KW		12192	2662	2662
Promoter Recycle Water Pump				1429	
Dryer Regen Heater	KW	1330	190	190	170
Other Loads	KW		22	13	9
MDEA AGR	KW		47	47	55
CO2 Removal Plant Loads		35766	57446	44265	29648
Cooling Tower Fans	KW	1214	1150	1198	1141
Balance of Plant	KW	53680	53680	53680	53680
Total Auxiliary loads	KW	90660	112276	99143	84469
OnStream Factor		0.8	0.8	0.8	0.8
Net Power	KW	384029	377027	398250	411710
Cost of Electricity	Cents/kWhr	6.24	6.65	6.13	5.75

Table 1-2 Equipment Cost Estimates

CO_2 Removal and Compression	Selexol	SIMTECHE		
		2-Stage Promoter	*1-Step 3000 Psia Reactor*	*Once-Thru*
CO_2 Separation Ratio	90%	90%	90%	68%
Total Installed Cost, $ 2Q 2005				
Vessels, Exchangers, Pumps	$ 39,987,946	$ 48,049,292	$ 64,933,505	$ 32,227,217
Refrigeration Compression	$ 3,046,633	$ 19,773,269	$ 16,174,463	$ 12,999,977
CO_2 + Other Compression	$ 17,817,452	$ 8,323,126	$ 17,749,899	$ 5,795,199
MDEA AGR	Not Required	$ 555,000	$ 555,000	$ 584,000
Column Packing, Initial Fill	$ 1,069,338	$ 67,874	$ 59,385	$ 24,521
Total CO2 Removal and Compression	**$61,921,368**	**$76,768,561**	**$99,472,251**	**$51,630,914**

Table 1-3 Power Production Summary

Power Production Summary	No Capture	Selexol	SIMTECHE		
			2-Stage Promoter	*1-Step 3000 Psia Reactor*	*Once-Thru*
CO_2 Separation Ratio	0%	90%	90%	91%	68%
Gross Plant Power, kW_e	474,040	474,689	497,392	489,303	496,179
Auxilliary Power Loads					
CO_2 Capture and Compression		35,766	44,265	57,446	29,648
Power Plant	49,500	54,894	54,878	54,830	54,821
Net Plant Power, kW_e	424,540	384,029	398,249	377,027	411,710
Net Efficiency, %HHV	43.0	35.2	36.5	34.6	37.7
Heat Rate, BTU/kWhr	7,936	9,693	9,347	9,872.8	9,041
Capture Parasitic Load, %		7.5%	8.9%	11.7%	6.0%

Table 1-4 Carbon Control Cost

Carbon Control Costs	No Capture	Selexol	SIMTECHE		
			2-Stage Promoter	*1-Step 3000 Psia Reactor*	*Once-Thru*
CO_2 Separation Ratio	0%	90%	90%	91%	68%
Onstream Factor	80%	80%	80%	80%	80%
Coal Feed (as received), Short Tons per Year	1,009,152	1,118,068	1,118,068	1,118,068	1,118,068
Carbon Dioxide Captured, STPY		2,288,719	2,278,145	2,297,500	1,724,192
Carbon Dioxide Emitted, STPY	2,353,363	237,382	247,995	228,640	801,947
Cost of Production, $ per Year					
Capital Charge	$ 87,113,705	$ 112,484,409	$ 115,290,529	$ 119,581,526	$ 110,539,513
Coal Cost	$ 29,265,408	$ 32,423,972	$ 32,423,972	$ 32,423,972	$ 32,423,972
Chemicals/Consumables	$ 2,271,894	$ 2,353,786	$ 2,285,469	$ 2,282,874	$ 2,276,798
Maintenance Cost, Materials&Labor	$ 9,218,382	$ 11,903,112	$ 12,200,056	$ 12,654,130	$ 11,697,303
Operating Labor and Administration	$ 8,688,493	$ 8,830,182	$ 8,830,182	$ 8,830,182	$ 8,830,182
Total Cost of Production, $ per year	**$ 136,557,882**	**$ 167,995,461**	**$ 171,030,207**	**$ 175,772,684**	**$ 165,767,768**
Net Power Produced, kW_e	424,540	384,029	398,249	377,027	411,710
Cost of Electricity, cents per kWhr	4.59	6.24	6.13	6.65	5.75
%Increase in Cost of Electricity		36%	34%	45%	25%
Avoided Capture Cost, $ per Ton of CO2	**$ -**	**$ 25**	**$ 24**	**$ 32**	**$ 25**
Avoided Capture Cost, $ per Ton of Carbon	**$ -**	**$ 6.9**	**$ 6.4**	**$ 8.6**	**$ 6.8**
Cost of Shift,$ per Ton of Carbon		$ 6.26	$ 5.93	$ 6.05	$ 7.80

Table 1-5 Utilities Summary

Item No	Item Name	Load BHP		Elect. Power	Cooling Water		Refrigeration	
		Norm.	Max (3).	KW	CW, MMbtu/hr	C.W. circ. GPM (2)	NH3, MMbtu/hr	Tons of Refrig.
	Exchangers							
E-100	1st Stg Syngas Compressor CW Cooler				15.9	2,120		
E-101	Syngas Chiller						15.50	1,292
E-102A	Hydrate Flash Reactor A							
E-102B	Hydrate Flash Reactor B							
E-103	Flash Gas Cooler				4.03	537		
E-104	Recycle Water Chiller						75.5	6,292
E-106	Syngas Cooler				40.14	5,352		
E-107A	Acid Gas Compr 1st Intercooler				0.11	11		
E-107B	Acid Gas Compr 2nd Intercooler				0.05	5		
E-107C	Acid Gas Compr 3rd Intercooler				0.05	5		
E-110	CO2 Sequestration Comp Aftercooler				47.66	6,354		
E-201	Ammonia Condenser				80.26	8,776		
	Hydrate Formation Reactors							
	Hydrate Formation Reactor						434.76	36,230
	Compressors							
K-100	Flash Gas Compressor	1,666		1,308	0.08	11		
K-101	NH_3 Compressor - Stage 1	25,051		19,672	1.28	170		
K-102	NH_3 Compressor - Stage 2	3,653		2,869	0.19	25		
K-103	Acid Gas Compressor	54		45	0.00	0		
K-104A	1st Stg Syngas Compressor	13,646		10,716	0.69	93		
K-104B	2nd Stg Syngas Compressor	13,299		10,443	0.68	90		
K-105	1st Stg Syngas Expander	(5,681)		(4,026)	0.29	39		
K-106	2nd Stg Syngas Expander	(4,989)		(3,536)	0.25	34		
K-107	CO2 Sequestration Compressor	8,971		7,044	0.46	61		
	Pumps							
	Chilled Water Recycle Pumps	14,710		12,193				
	Cooling Water Pumps	553		459				
	Lube Oil Pumps	24		20				
	Demineralized Water Pumps	0.03		0.02				
	TOTAL	70,956		57,206	192.11	23,683	525.76	43,813

NOTES:

1 All Figures shown above represent normal utility usage requirements except:

2 Cooling water supply temperature is 80 F. Makeup water to cooling tower is not shown

3 Utility consumption for max. load conditions is not shown.

Table 1-6 Equipment List

Equipment	No. ea.	Size/Duty	Material Tube/Shell	Temperature (°F) Tube/Shell	Pressure (psig) Tube/Shell
Reactors					
Hydrate Formation Reactors	10	435 MM Btu/hr total	304 SS/CS	200/200	3150/235
Hydrate Flash Reactor A	8	476.2 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/680
Hydrate Flash Reactor B	1	40.5 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/315
Exchangers					
1st Stg Syngas CW Cooler	1	15.9 MM Btu/hr	CS/CS	285/250	3200/235
Syngas Chiller	1	15.5 MM Btu/hr	304 SS/CS	200/200	1,100/235
Recycle Water Chiller	2	73.6. MM Btu/hr	304 SS/CS	200/200	300/235
Flash Gas Cooler	1	4.0 MM Btu/hr	CS/CS	285/200	680/50
Syngas/Product Gas Exch.	1	8.6 MM Btu/hr	CS /304 SS	250/250	3240/3030
2nd Stg Syngas CW Cooler	1	40.1 MM Btu/hr	304 SS/CS	250/250	75/3230
CO_2 Sequestration Cooler	1	47.7 MM Btu/hr	CS/CS	375/200	2,400/50
Ammonia Condenser	4	80.3 MM Btu/hr	CS/CS	200/360	500/235
Acid Gas Compressor 1st Stage Intercooler	1	0.11 MM Btu/hr	CS/304 SS (Clad)	200/350	50/50
Acid Gas Compressor 2nd Stage Intercooler	1	0.05 MM Btu/hr	CS/304 SS (Clad)	200/350	50/110
Acid Gas Compressor 3rd Stage Intercooler	1	0.05 MM Btu/hr	CS/304 SS (Clad)	200/350	50/320
Syngas/2nd Stg Expander Gas Exch	1	11.7 MM Btu/hr	CS/304 SS (Clad)	250/250	1900/1700
Syngas/1st Stg Expander Gas Exch	1	11.7 MM Btu/hr	304 SS/CS	250/250	30201875
Regeneration Heater-Electric	1	190 kW	CS/CS	400	15
Vessels					
Syngas KO Drum	1	8'ID x 10' T-T	304 SS Clad	250	1,055
Slurry/Gas Separator	1	8.5'ID x 35' T-T	304 SS Clad	250	3030
Flash Gas Compr KO Drum	1	4.0'ID x 7.5' T-T	304 SS Clad	120	300
Acid Gas Compr KO Drum	1	1'ID x 5' T-T	304 SS Clad	160	50
Acid Gas 1st Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	100
Acid Gas 2nd Stg Compr KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	230
Acid Gas Product KO Drum	1	1'ID x 4' T-T	304 SS Clad	160	670
Mol Sieve Dryer	2	7'ID x 16' T-T	304 SS Clad	450	670
Ammonia Separator	1	4.5'ID x 15' T-T	CS	200	235
Ammonia Surge Drum	1	12'ID x 37' T-T	CS	200	235
Ammonia Compr KO Drum	1	15'ID x45' T-T	CS	200	235
2nd Stg Syngas Compressor KO Drum	1	7'ID x 7' T-T	304 SS Clad	250	1,900
Compressors					
Flash Gas Compressor	1	2,000 HP	CS	350	625
CO2 Sequestration Compressor	1	10,000 HP	CS	350	2400
Acid Gas Compressor	1	75 HP	CS	350	610
NH3 Refrig 1st Stage Compressor	1	27,500 HP	CS	200	136
NH3 Refrig 2nd Stage Compressor	1	4,000 HP	CS	350	245

1st Stg Syngas Compressor	1	15,000 HP	CS	350	1900
2nd Stg Syngas Compressor	1	15,000 HP	CS	350	3200
Equipment	No. ea.	Size/Duty	Material	Temperature (°F)	Pressure (psig)
1st Stg Syngas Expander	1	5,400 HP	CS	250	3030
2nd Stg Syngas Expander	1	4,740 HP	CS	250	1900
Pumps					
Chilled Water Recycle Pumps (3-50% pumps)	3	3700 GPM/ 8,200 HP each	304 SS	200	1100
Cooling Water Pumps (3-50% pumps)	3	12,000 GPM/350 HP each	CS	200	100
Demineralized Water Pumps (2-100% pumps)	2	0.14 GPM/0.03 HP each	CS		
Lube Oil Pumps (3-50% pumps)	3	975 GPM/24 HP each	CS		

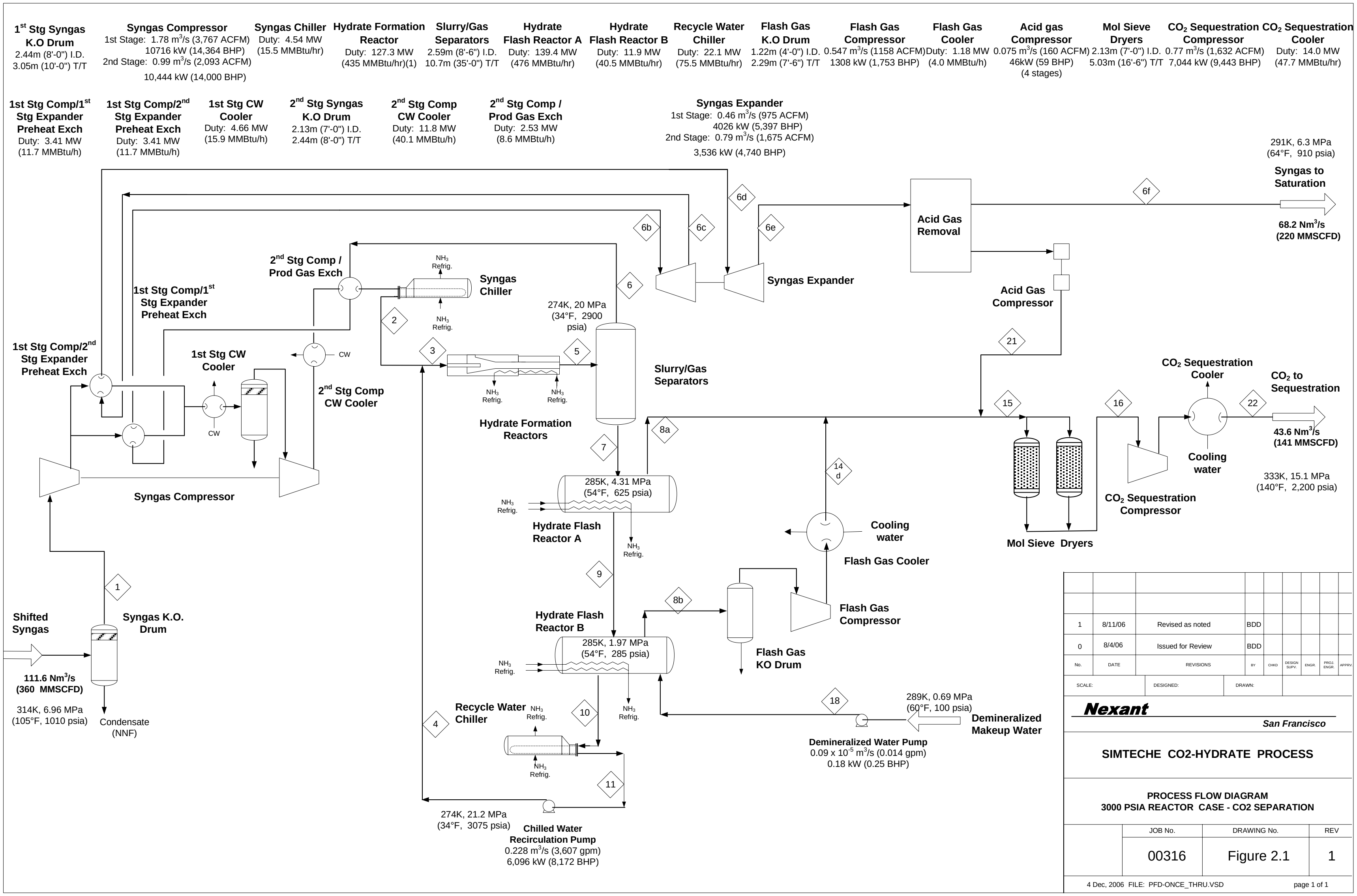
Table 1-7 Material Balance

ONCE THROUGH CASE- 3000 PSIA HYDRATE FORMATION REACTOR- HEAT & MATERIAL BALANCE - LANL HYDRATE CORRELATION

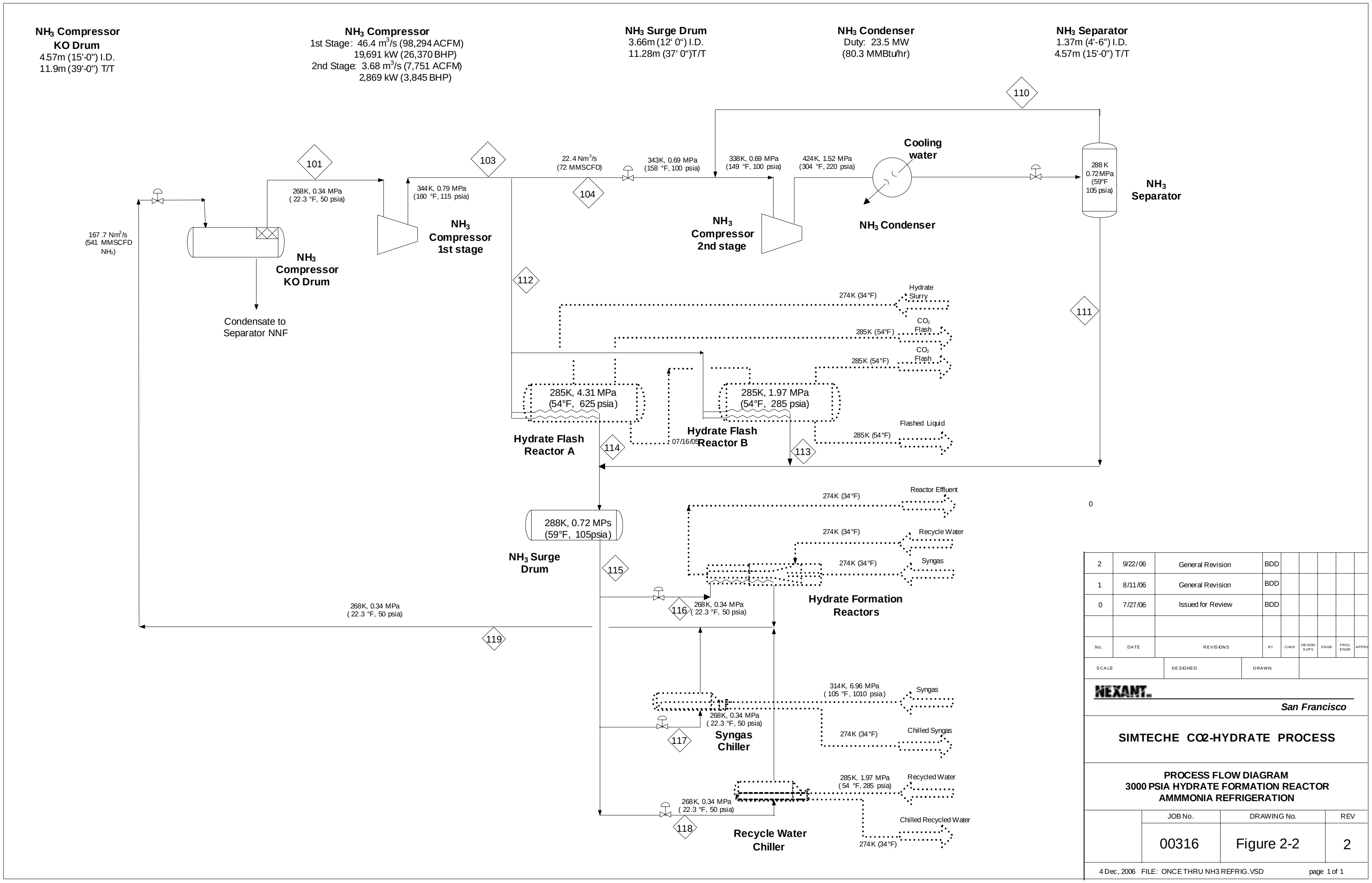
Coal	**Illinois No. 6**	**FLASH A PRESSURE, PSIA**	**625**			Treated Gas to GT H2S Spec =	40 ppmV
Wt% Sulfur	**2.82 wt% dry(2.51wt% S As Received)**	**FLASH B PRESSURE, PSIA**	**285**	**SR**	**0.906**		

Stream Numbers		Strm 1	Strm 1	Strm 1A	Strm 1A	Strm 1B	Strm 1B	Strm 2	Strm 2	Strm 3	Strm 3	Strm 4	Strm 4	Strm 5	Strm 5	Strm 6	Strm 6	Strm 7	Strm 7	Strm 8A	Strm 8A	Strm 8B	Strm 8B	Strm 9	Strm 9	Strm 10	Strm 10
		Sat'd Syngas to KO drum	Sat'd Syngas to KO drum	Sat'd Syngas to chiller	Sat'd Syngas to chiller	KO drum Condensate	KO drum Condensate	Chilled Syngas	Chilled Syngas	Hform Rx Feed	Hform Rx Feed	Nucleated Water	Nucleated Water	Reactor Effluent	Reactor Effluent	Treated Syngas to AGR	Treated Syngas to AGR	Slurry fr HFR-A	Slurry fr HFR-A	CO_2 Vapor fr HFR-A	CO_2 Vapor fr HFR-A	CO_2 Vapor fr HFR-B	CO_2 Vapor fr HFR-B	Water fr fr HFR-A	Water fr fr HFR-A	Water fr fr HFR-B	Water fr fr HFR-B
	Mol Wt	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h
H_2	2.02	21,418.72	43,176	21,418.72	43,176	-	-	21,418.72	43,176	21,418.77	43,176	0.05	0	21,418.77	43,176	21,129.17	42,592	289.60	584	286.15	577	3.39	7	3.44	7	0.05	0
CO_2	44.01	16,381.13	720,930	16,381.13	720,930	-	-	16,381.13	720,930	19,768.94	870,027	3,387.81	149,097	3,162.33	139,174	1,533.60	67,493	1,628.73	71,680	10,800.44	475,325	4,047.09	178,112	7,434.90	327,208	3,387.81	149,097
H_2S	34.08	248.44	8,466	248.44	8,466	-	-	248.44	8,466	468.77	15,974	220.33	7,508	30.62	1,044	8.06	275	22.57	769	151.56	5,164	88.83	3,027	309.16	10,535	220.33	7,508
H_2O Vap	18.02	1.27	23	1.27	23	-	-	1.27	23	1.27	23	-	-	1.27	23	0.78	14	-	-	3.57	64	2.94	53	-	-	-	-
Ar	39.95	292.72	11,694	292.72	11,694	-	-	292.72	11,694	292.72	11,694	0.00	0	292.72	11,694	288.47	11,524	4.25	170	4.20	168	0.05	2	0.05	2	0.00	0
N2	28.01	289.32	8,105	289.32	8,105	-	-	289.32	8,105	289.32	8,105	0.00	0	289.32	8,105	285.12	7,987	4.20	118	4.16	116	0.05	1	0.05	1	0.00	0
CO	28.01	382.31	10,709	382.31	10,709	-	-	382.31	10,709	382.31	10,709	0.00	0	382.31	10,709	374.02	10,476	8.29	232	8.15	228	0.13	4	0.14	4	0.00	0
CH4	16.04	537.79	8,628	537.79	8,628	-	-	537.79	8,628	537.80	8,628	0.01	0	537.80	8,628	519.84	8,340	17.97	288	17.53	281	0.42	7	0.43	7	0.01	0
COS	60.07	0.04	2	0.04	2	-	-	0.04	2	0.06	4	0.02	1	0.06	4	0.02	1	0.04	3	0.02	1	0.01	1	0.03	2	0.02	1
NH3	17.03	3.40	58	3.40	58	-	-	3.40	58	415.86	7,082	412.46	7,024	415.86	7,082	1.05	18	414.81	7,064	1.29	22	1.06	18	413.52	7,042	412.46	7,024
H_2O Liq	18.02	-	-	-	-	-	-	-	-	198,045.87	3,567,836	198,045.87	3,567,836	95,777.34	1,725,448	-	-	95,777.82	1,725,457	-	-	-	-	198,042.78	3,567,780	198,045.87	3,567,836
CO_2 Hydrate	152.10	-	-	-	-	-	-	-	-	-	-	-	-	16,606.61	2,525,882	-	-	16,606.61	2,525,882	-	-	-	-	-	-	-	-
H_2S Hydrate	142.17	-	-	-	-	-	-	-	-	-	-	-	-	438.15	62,290	-	-	438.15	62,290	-	-	-	-	-	-	-	-
Total		39,555.14	811,790	39,555.14	811,790	-	-	39,555.14	811,790	241,621.70	4,543,256	202,066.56	3,731,467	139,353.16	4,543,256	24,140.13	148,720	115,213.03	4,394,536	11,277.07	481,948	4,143.96	181,230	206,204.50	3,912,588	202,066.56	3,731,467
Temperature, F		105		105		105		34		34		34		34		34		34		54		54		54		54	
Pressure, psia		1,010		1,010		1,010		3,000		3,000		3,050		2,910		2,900		2,900		625		285		625		285	

Stream Numbers		Strm 11	Strm 11	Strm 14A	Strm 14A	Strm 14B	Strm 14B	Strm 14C	Strm 14C	Strm 14D	Strm 14D	5	15	Strm 15	Strm 15	Strm 16	Strm 16	Strm 17	Strm 17	Strm 18	Strm 18	Strm 6A	Strm 6A	Strm 21	Strm 21	Strm 22	Strm 22
		Chilled Recycle Water	Chilled Recycle Water	FlashGas Compr KO Liq	FlashGas Compr KO Liq	FlashGas Compr Feed	FlashGas Compr Feed	FlashGas Compr Discharge	FlashGas Compr Discharge	Cooled FG Compr Discharge	Cooled FG Compr Discharge	Total CO_2 to Dryer	Total CO_2 to Dryer	CO2 Seq Compr Discharge	CO2 Seq Compr Discharge	Cooled CO2 to Seq	Cooled CO_2 to Seq	Total Condensate	Total Condensate	Demin Water Makeup	Demin Water Makeup	Treated Gas to GT	Treated Gas to GT	Acid Gas fr AGR	Acid Gas fr AGR	Total CO2 to Dryers	Total CO2 to Dryers
	Mol Wt	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h
H_2	2.02	0.05	0	-	-	3.39	7	3.39	7	3.39	7	289.55	584	289.55	584	289.55	584	-	-	-	-	21,129.17	42,592	-	-	289.55	584
CO_2	44.01	3,387.81	149,097	-	-	4,047.09	178,112	4,047.09	178,112	4,047.09	178,112	14,847.53	653,437	14,898.48	655,679	14,898.48	655,679	-	-	-	-	1,482.65	65,251	50.95	2,242	14,898.48	655,679
H_2S	34.08	220.33	7,508	-	-	88.83	3,027	88.83	3,027	88.83	3,027	240.38	8,191	265.22	9,038	265.22	9,038	-	-	-	-	(16.78)	(572)	24.84	846	265.22	9,038
H_2O Vap	18.02	-	-	-	-	2.94	53	2.94	53	2.94	53	6.51	117	-	-	-	-	-	-	-	-	33.56	605	11.89	214	18.40	331
Ar	39.95	0.00	0	-	-	0.05	2	0.05	2	0.05	2	4.25	170	4.25	170	4.25	170	-	-	-	-	288.47	11,524	-	-	4.25	170
N2	28.01	0.00	0	-	-	0.05	1	0.05	1	0.05	1	4.20	118	4.20	118	4.20	118	-	-	-	-	285.12	7,987	-	-	4.20	118
CO	28.01	0.00	0	-	-	0.13	4	0.13	4	0.13	4	8.29	232	8.29	232	8.29	232	-	-	-	-	374.02	10,476	-	-	8.29	232
CH4	16.04	0.01	0	-	-	0.42	7	0.42	7	0.42	7	17.95	288	17.95	288	17.95	288	-	-	-	-	519.84	8,340	-	-	17.95	288
COS	60.07	0.02	1	-	-	0.01	1	0.01	1	0.01	1	0.02	1	0.02	1	0.02	1	-	-	-	-	0.02	1	-	-	0.02	1
NH3	17.03	412.46	7,024	-	-	1.06	18	1.06	18	1.06	18	2.35	40	2.35	40	2.35	40	-	-	-	-	1.05	18	-	-	2.35	40
H_2O Liq	18.02	198,045.87	3,567,836	-	-	-	-	-	-	-	-	-	-	-	-	-	-	18.40	331	6.02	108	-	-	-	-	-	-
CO_2 Hydrate	152.10	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2S Hydrate	142.17	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total		202,066.56	3,731,467	-	-	4,143.96	181,230	4,143.96	181,230	4,143.96	181,230	15,421.03	663,178	15,490.31	666,150	15,490.31	666,150	18.40	331	6.02	108	24,097.12	146,222	87.68	3,303	15,508.71	666,481
Temperature, F		34		54		54		182		110		68		276		140		100		54		105		244		68	
Pressure, psia		280		285		285		630		625		625		2,210		2,200		625		295		922		625		625	



1st Stg Syngas K.O Drum
2.44m (8'-0") I.D.
3.05m (10'-0") T/T
Syngas Compressor
1st Stage: 1.78 m3/s (3,767 ACFM)
10716 kW (14,364 BHP)
2nd Stage: 0.99 m3/s (2,093 ACFM)
10,444 kW (14,000 BHP)
Syngas Chiller
Duty: 4.54 MW
(15.5 MMBtu/hr)
Hydrate Formation Reactor
Duty: 127.3 MW
(435 MMBtu/hr)(1)
Slurry/Gas Separators
2.59m (8'-6") I.D.
10.7m (35'-0") T/T
Hydrate Flash Reactor A
Duty: 139.4 MW
(476 MMBtu/hr)
Hydrate Flash Reactor B
Duty: 11.9 MW
(40.5 MMBtu/hr)
Recycle Water Chiller
Duty: 22.1 MW
(75.5 MMBtu/hr)
Flash Gas K.O Drum
1.22m (4'-0") I.D.
2.29m (7'-6") T/T
Flash Gas Compressor
0.547 m3/s (1158 ACFM)
1308 kW (1,753 BHP)
Flash Gas Cooler
Duty: 1.18 MW
(4.0 MMBtu/h)
Acid gas Compressor
0.075 m3/s (160 ACFM)
46kW (59 BHP)
(4 stages)
Mol Sieve Dryers
2.13m (7'-0") I.D.
5.03m (16'-6") T/T
CO2 Sequestration Compressor
0.77 m3/s (1,632 ACFM)
7,044 kW (9,443 BHP)
CO2 Sequestration Cooler
Duty: 14.0 MW
(47.7 MMBtu/hr)
1st Stg Comp/1st Stg Expander Preheat Exch
Duty: 3.41 MW
(11.7 MMBtu/h)
1st Stg Comp/2nd Stg Expander Preheat Exch
Duty: 3.41 MW
(11.7 MMBtu/h)
1st Stg CW Cooler
Duty: 4.66 MW
(15.9 MMBtu/h)
2nd Stg Syngas K.O Drum
2.13m (7'-0") I.D.
2.44m (8'-0") T/T
2nd Stg Comp CW Cooler
Duty: 11.8 MW
(40.1 MMBtu/h)
2nd Stg Comp / Prod Gas Exch
Duty: 2.53 MW
(8.6 MMBtu/h)
Syngas Expander
1st Stage: 0.46 m3/s (975 ACFM)
4026 kW (5,397 BHP)
2nd Stage: 0.79 m3/s (1,675 ACFM)
3,536 kW (4,740 BHP)
291K, 6.3 MPa
(64°F, 910 psia)
Syngas to Saturation
68.2 Nm3/s
(220 MMSCFD)
Acid Gas Removal
6d
6f
6b
6c
6e
2nd Stg Comp / Prod Gas Exch
NH3 Refrig.
Syngas Chiller
Syngas Expander
Acid Gas Compressor
1st Stg Comp/1st Stg Expander Preheat Exch
6
274K, 20 MPa
(34°F, 2900 psia)
2
1st Stg Comp/2nd Stg Expander Preheat Exch
1st Stg CW Cooler
CW
3
5
21
2nd Stg Comp CW Cooler
Slurry/Gas Separators
CO2 Sequestration Cooler
CO2 to Sequestration
Hydrate Formation Reactors
8a
15
16
22
43.6 Nm3/s
(141 MMSCFD)
7
Cooling water
14d
285K, 4.31 MPa
(54°F, 625 psia)
333K, 15.1 MPa
(140°F, 2,200 psia)
Syngas Compressor
CO2 Sequestration Compressor
Hydrate Flash Reactor A
Cooling water
Flash Gas Cooler
Mol Sieve Dryers
9
1
8b
Shifted Syngas
Syngas K.O. Drum
Hydrate Flash Reactor B
Flash Gas Compressor
285K, 1.97 MPa
(54°F, 285 psia)
Flash Gas KO Drum
111.6 Nm3/s
(360 MMSCFD)
314K, 6.96 MPa
(105°F, 1010 psia)
Condensate (NNF)
Recycle Water Chiller
10
4
18
289K, 0.69 MPa
(60°F, 100 psia)
Demineralized Makeup Water
Demineralized Water Pump
0.09 x 10-5 m3/s (0.014 gpm)
0.18 kW (0.25 BHP)
11
274K, 21.2 MPa
(34°F, 3075 psia)
Chilled Water Recirculation Pump
0.228 m3/s (3,607 gpm)
6,096 kW (8,172 BHP)
1 8/11/06 Revised as noted BDD
0 8/4/06 Issued for Review BDD
No. DATE REVISIONS BY CHKD DESIGN SUPV. ENGR. PROJ. ENGR. APPRV.
SCALE: DESIGNED: DRAWN:
Nexant
San Francisco
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
3000 PSIA REACTOR CASE - CO2 SEPARATION
JOB No. DRAWING No. REV
00316 Figure 2.1 1
4 Dec, 2006 FILE: PFD-ONCE_THRU.VSD page 1 of 1


NH3 Compressor KO Drum
4.57m (15'-0") I.D.
11.9m (39'-0") T/T
NH3 Compressor
1st Stage: 46.4 m3/s (98,294 ACFM)
19,691 kW (26,370 BHP)
2nd Stage: 3.68 m3/s (7,751 ACFM)
2,869 kW (3,845 BHP)
NH3 Surge Drum
3.66m (12' 0") I.D.
11.28m (37' 0")T/T
NH3 Condenser
Duty: 23.5 MW
(80.3 MMBtu/hr)
NH3 Separator
1.37m (4'-6") I.D.
4.57m (15'-0") T/T
110
101
103
104
Cooling water
268K, 0.34 MPa
(22.3 °F, 50 psia)
344K, 0.79 MPa
(160 °F, 115 psia)
22.4 Nm3/s
(72 MMSCFD)
343K, 0.69 MPa
(158 °F, 100 psia)
338K, 0.69 MPa
(149 °F, 100 psia)
424K, 1.52 MPa
(304 °F, 220 psia)
288 K
0.72 MPa
(59°F
105 psia)
NH3 Separator
167.7 Nm3/s
(541 MMSCFD
NH3)
NH3 Compressor KO Drum
NH3 Compressor 1st stage
NH3 Compressor 2nd stage
NH3 Condenser
Condensate to Separator NNF
112
111
274K (34°F)
Hydrate Slurry
285K (54°F)
CO2 Flash
285K (54°F)
CO2 Flash
285K, 4.31 MPa
(54°F, 625 psia)
285K, 1.97 MPa
(54°F, 285 psia)
Flashed Liquid
285K (54°F)
Hydrate Flash Reactor A
Hydrate Flash Reactor B
114
113
07/16/09
274K (34°F)
Reactor Effluent
0
288K, 0.72 MPs
(59°F, 105psia)
NH3 Surge Drum
274K (34°F)
Recycle Water
274K (34°F)
Syngas
115
116
268K, 0.34 MPa
(22.3 °F, 50 psia)
Hydrate Formation Reactors
268K, 0.34 MPa
(22.3 °F, 50 psia)
119
314K, 6.96 MPa
(105 °F, 1010 psia)
Syngas
268K, 0.34 MPa
(22.3 °F, 50 psia)
274K (34°F)
Chilled Syngas
Syngas Chiller
117
285K, 1.97 MPa
(54 °F, 285 psia)
Recycled Water
268K, 0.34 MPa
(22.3 °F, 50 psia)
Chilled Recycled Water
118
Recycle Water Chiller
274K (34°F)
2 | 9/22/06 | General Revision | BDD
1 | 8/11/06 | General Revision | BDD
0 | 7/27/06 | Issued for Review | BDD
No. | DATE | REVISIONS | BY | CHKD | DESIGN SUPV | ENGR | PROJ ENGR | APPR
SCALE | DESIGNED | DRAWN
NEXANT
San Francisco
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
3000 PSIA HYDRATE FORMATION REACTOR
AMMMONIA REFRIGERATION
JOB No. | DRAWING No. | REV
00316 | Figure 2-2 | 2
4 Dec, 2006 FILE: ONCE THRU NH3 REFRIG.VSD page 1 of 1

Task 5 – 1800 Psia Hydrate Formation Reactor Operating Pressure

Objective

The results of Task 4 (3000 psia hydrate formation reactor operating pressure case) indicates that considerably higher capital and operating costs will be incurred if the Simteche process is to recover 90% of the CO_2 in a single pass without the addition of a promoter. An optimum recovery for the Simteche process is probably less than 90%. A logical pressure to operate the Hydrate Formation reactor is probably around 1800 psia which can be achieved with a single stage syngas compressor from either a low pressure (500 psig) gasifier or from a high pressure (1050 psig) gasifier. Experimental observations at LANL showed that hydrate formation can be greatly accelerated by increasing the CO_2 partial pressure and preliminary Simteche calculations also showed that an 84% CO_2 recovery can be achieved at hydrate reactor operating conditions of 1,800 psia and 34 ^oF.

The objective of this task is to revise the process scheme to allow the hydrate formation reactor to be operated at an intermediate pressure of 1800 psia and 34^oF and re-assess the overall process techno-economics.

Summary

The capital cost and energy consumption of the CO_2 hydrate process increase rapidly as the hydrate formation reactor operating pressure is increased in order to achieve higher levels of CO_2 recovery, as shown in Figure 1. Capital cost and energy consumption increase are mainly due to the following:

- Higher cost for equipment associated with high pressure operation.
- Additional equipment requirement such as two stages of high pressure compression and expansion are required.
- Significant energy loads for the recycle water pumps.
- CO_2 must be recovered at the same low pressures as the base case, 1000 psia equates to no savings in CO_2 compression costs compared to the base case.

A techno-economic analysis was performed comparing process schemes of different hydrate formation reactor pressure at different levels of CO_2 recovery – i.e., 1000, 1800 and 3000 psia using the once-through Simteche CO_2 hydrate design. The analysis was based on a fully integrated IGCC (EPRI prototype) plant design with CO_2 capture.

Table 1-1 summarizes the results between the various cases, with the estimated plant capital costs, in-plant power consumption, and cost of electricity production plotted against percentage of CO_2 recovery resulted from increasing the hydrate formation pressure, as shown in Figures 1 & 2. As can be seen, capital costs, power consumption and cost of electricity production increase with increasing level of CO_2 recovery, and these increases take on a drastic rise at CO_2 recovery beyond about 85%. Selexol data at 90% CO_2 recovery were also included in Figures 1 & 2 for comparison.

1800 Psia Hydrate Formation Reactor Case – Process Scheme

The 1800 psia hydrate formation reactor scheme is shown in Figures 2-1 and 2-2. In this medium pressure variation of the once-through Simteche CO_2 recovery process, compared to the 3000 psia case analyzed in Task 4, the syngas is compressed and intercooled in a single stage compressor to 1850 psia This is followed with cooling by exchange with syngas product, cooling water, and ammonia to 34^oF (see Fig 2.1). It is then mixed with recycled chilled water and sent to the hydrate formation reactors. Here the mixture is chilled by exchange with ammonia and the resultant CO_2 and H_2S hydrates are formed. The reactor effluent is sent to a slurry/gas separator which operates at 1700 psia and 34^oF and where the phases are separated. The gas stream is the heated first by exchange with syngas feed and then by exchange with the first stage syngas compressor discharge to preheat it to 195 ^oF before being sent to single stage turbo expander. This preheat is sufficient to avoid any ice or other solids formation in the expander and downstream equipment. The effluent from the expander exhausts at 922 psia and 110 ^oF and where it is sent to an acid gas removal unit which removes H_2S to meet the syngas product spec of 40 ppm sulfur for the gas turbine. The final product is then sent to syngas humidification. A small waste gas stream from the acid gas removal unit containing H_2S and a small amount of CO_2 is compressed to 630 psia.

The liquid slurry from the slurry/gas separator is letdown in pressure to 625 psia and the slurry is heated to 54 ^oF in hydrate flash reactor A. Here the Hydrate is disassociated to CO_2, H_2S and water and 60% of the CO_2 is flashed off and sent to the mol sieve gas dehydration unit. The heating is accomplished by condensing part of the 1st stage NH3 refrigeration compressor effluent. The remaining liquid containing water, CO_2 and H_2S is letdown in pressure to 285 psia and sent to hydrate flash reactor B. Here the mixture is heated again and additional CO_2 and H2S are flashed off. This heating is also accomplished by condensing part of the 1st stage NH_3 refrigeration compressor effluent. This off gas is compressed in a flash gas compressor to 630 psia and mixed with the off gas from hydrate flash reactor A and the waste gas from the acid gas removal unit. This mixture is then sent to a mol sieve gas dehydration unit to meet the product CO_2 dryness specification. The dried CO_2 is then compressed in the CO_2 sequestration compressor and cooled and sent offsite for sequestration.

The liquid from the hydrate flash reactor B containing mostly water and dissolved CO_2 and H_2S is then chilled to 34 ^oF and pumped back to the inlet of the Hydrate Formation reactor.

Techno-economic Analysis - Methodology

An existing spreadsheet model, utilized in designing the cases included in the Engineering Analysis Report of February 2006, was used for to develop the material balances. It incorporated the Henry's law coefficients to handle the flashes with the hydrate present. A simulation model was built using Hysys to develop the heat and mass balance and facilitate sizing and costing the plant. The Hysys model was modified to handle the equilibrium flashes using the same Henry's law constants that were employed in the spreadsheet. The simulation model was unable to recognize hydrate formation or hydrate disassociation. The heat balance was adjusted manually with the heats of hydrate formation as determined below. The heat balance around the hydrate formation reactor and the hydrate heater were handled separately using Hysys heat capacities and heats of formation of the hydrates. The total CO_2 heats of hydrate formation (vapor phase to solid hydrate) was assumed to be 62 kJ/gmol[1]. Heat of hydrate formation of CO_2 and H_2S (solid hydrate to vapor phase) were assumed to be the same at 1800 psia as at 1000 psia. The H_2S heats of hydrate formation were assumed to be the same as CO_2 on a molar basis since they had similar hydration numbers and any deviation was expected to be small since the major contributor is the heat of fusion of water. In addition the heat of hydration of H_2S represents only 2% of the hydrate formation reactor duty. These heats of hydrate formation will need to be verified by experimental data.

The property packages used in Hysys are known to be accurate in the 1000 to 3000 psia conditions of the syngas compression and expansion.

Because the high pressure flash takes place at 1700 psia well above the syngas product delivery pressure of 900 psia , additional energy can be recovered from this stream by means of a gas turbo expander to reduce the syngas pressure to 920 psia. However expanders can result in low discharge temperatures where ice and hydrates can form. To avoid the potential for low temperatures the expander feed gas is preheated with exchange with the syngas compressor discharge streams. This has the following advantages:

- Reduces the cooling water requirement for the syngas compressors
- Increases the power attainable from the gas turbines
- Eliminates hydrate formation and localized freezing in the downstream acid gas removal unit

The 1800 psia operating conditions case along with its comparison with the other cases are presented in the following:

Table 1-1 - Economic Comparative Analysis
Table 1-2 - Equipment Cost Estimates
Table 1-3 - Power Production Summary
Table 1-4 - Carbon Control Cost
Table 1.5 - Utility summary
Table 1.6 - Equipment list
Table 1.7- Material Balance
Figure 1 - CO_2 Recovery vs Capital Cost and Energy Consumption

[1] "Syngas Upgrading – A Low Temperature Approach, Phase 1 Report, June 2002" by Los Alamos National Laboratory, Simteche, and Nexant

Figure 2 - CO_2 Recovery vs Cost of Electricity
Figure 2.1 - Simplified PFD CO_2 Recovery Section
Figure 2.2 - Simplified PFD NH_3 Refrigeration Section

Results

In trying to increase the CO_2 recovery from 68 to 90% it becomes ever more difficult to get the incremental CO2 recovery. As can be seen in Table 1-1 and Figure 1, to increase the CO_2 recovery from 68 to 90 % requires a doubling of the energy consumption for the CO_2 recovery plant from 30 to 57 MW. Thus a 33% increase in CO_2 recovery results in an 93% increase in power consumption. Whereas increasing the recovery from 68 to 83.5 % increases the energy consumption by 43% while increasing the CO_2 recovery by 22% The refrigeration load increased slightly less than the increase in CO_2 removal, but there was an additional 11 MW required for Syngas compression although a credit of 4.5 MW was gained by utilizing the turbo expanders to recover energy from the high pressure gas. However there was an incremental 3.7 MW required for the recycle water pump. In this task the Syngas is compressed up to twice the pressure that is required to deliver it to the fuel gas saturation section.

In Table 1.2 and Figure 1 shows that going from 1000 to 3000 psia operating conditions doubles the capital cost while only increasing the CO_2 recovery by 33% from 68 to 90%. Whereas going to from 1000 to 1800 psia increases the capital cost by 52% for a 22% increase the CO_2 recovery.

Table 1.4 and Figure 2 show that going from 1000 to 3000 psia operating conditions increases the cost of electricity by 16% while only increasing the CO_2 recovery by 33%, whereas going to from 1000 to 1800 psia increases the COE 7% for a 22% increase in the CO_2 recovery. Table 1.4 also shows that the avoided capture cost of CO_2 also progressively increases as the CO_2 recovery exceeds 84% going from 25 to 32 $/Ton CO_2.

In going to an intermediate level such as 84%, the incremental cost isn't quite as severe as going to 90%. Figure 2 shows the cost of electricity at the 3 different CO_2 recovery levels. As can be seen the cost of electricity progressively increases as the CO_2 recovery exceeds 80% such that the optimum point is something less than 1800 psia. The Selexol option still achieves better results than the once-thru Simteche process for a given CO2 recovery level.

Table 1-1 Economic Comparative Analysis

CO2 Recovery		**90**	**83.5**	**68**	**90**
Case		**Once-Thru 3000 Psia Hydrate Form Reactor**	**Once-Thru 1800 Psia Hydrate Form Reactor**	**Once-Thru 1000 Psia Hydrate Formation Reactor**	**Selexol**
Capital Cost	MM$	99.47	78.68	51.63	61.92
Power Revenues					
Gas Turbine Power	KW	341577	342642	342295	333720
Steam Turbine Power	KW	145814	148634	150314	139294
Generator Loss	KW	(7250)	(7272)	(7265)	(7330)
Turboset Power	KW	480141	484004	485344	465684
Fuel Gas Expander Power	KW	9162	9402	10835	9005
Total Gross Power Generated	KW	489303	493406	496179	474689
Auxiliary Loads	KW				
1st Stg NH3 Compressors	KW	19672	18280	16819	3787
2nd Stg NH3 Compressors	KW	2869	2560	3139	
Acid Gas Compressors	KW	46	76	152	
Flash Gas Compressors	KW	1308	1235	1040	9726
1st stg Syngas Compressor	KW	10716	11163		
2nd stg Syngas Compressor	KW	10443			
CO2 Sequestration Compressor	KW	7044	6537	5289	15380
1st Stg Product Gas Expander	KW	(4026)	(4468)		
2nd Stg Product Gas Expander	KW	(3536)			
Solvent Pumps	KW				5042
Cooling Water Pumps	KW	459	331	313	501
Recycle Water Pump	KW	12192	6439	2662	
Dryer Regen Heater	KW	190	180	170	1330
Other Loads	KW	22	17	9	
MDEA AGR	KW	47	52	55	0
CO2 Removal Plant Loads	KW	57446	42402	29648	35766
Cooling Tower Fans	KW	1150	1116	1141	1162
Balance of Plant	KW	53680	53680	53680	53680
Total Auxiliary loads	KW	112276	97198	84469	90608
OnStream Factor		0.8	0.8	0.8	0.8
Net Power	KW	377027	396208	411710	384081
Cost of Electricity	Cents/kWhr	6.24	6.65	6.17	5.75

Table 1-2 Equipment Cost Estimates

CO_2 Removal and Compression	Selexol	Simteche		
		1-Step 3000 Psia Reactor	*1-Step 1800 Psia Reactor*	*Once Thru 1000 Psia Reactor*
CO_2 Separation Ratio	90%	90.6%	84%	68%
Total Installed Cost, $ 2Q 2005				
Vessels, Exchangers, Pumps	$ 39,987,946	$ 64,933,505	$ 49,709,257	$ 32,227,217
Refrigeration Compression	$ 3,046,633	$ 16,174,463	$ 15,916,927	$ 12,999,977
CO_2 + Other Compression	$ 17,817,452	$ 17,749,899	$ 12,436,764	$ 5,795,199
MDEA AGR	Not Required	$ 555,000	$ 555,000	$ 584,000
Column Packing, Initial Fill	$ 1,069,338	$ 59,385	$ 59,385	$ 24,521
Total CO2 Removal and Compression	**$61,921,368**	**$99,472,251**	**$78,677,333**	**$51,630,914**

Table 1-3 Power Production Summary

Power Production Summary	No Capture	Selexol	Simteche		
			1-Step 3000 Psia Reactor	*1-Step 1800 Psia Reactor*	*Once Thru 1000 Psia Reactor*
CO_2 Separation Ratio	0%	90%	90.6%	84%	68%
Gross Plant Power, kW_e	474,040	474,689	489,303	493,406	496,179
Auxilliary Power Loads					
CO_2 Capture and Compression		35,766	57,446	42,402	29,648
Power Plant	49,500	54,894	54,830	54,796	54,821
Net Plant Power, kW_e	424,540	384,029	377,027	396,208	411,710
Net Efficiency, %HHV	43.0	35.2	34.6	36.3	37.7
Heat Rate, BTU/kWhr	7,936	9,693	9,872.8	9,394.9	9,041.1
Capture Parasitic Load, %		7.5%	11.7%	8.6%	6.0%

Table 1- 4 Carbon Control Cost

Carbon Control Costs	No Capture	Selexol	Simteche		
			1-Step 3000 Psia Reactor	*1-Step 1800 Psia Reactor*	*Once Thru 1000 Psia Reactor*
CO_2 Separation Ratio	0%	90%	90.6%	83.5%	68.0%
Onstream Factor	80%	80%	80%	80%	80%
Coal Feed (as received), Short Tons per Year	1,009,152	1,118,068	1,118,068	1,118,068	1,118,068
Carbon Dioxide Captured, STPY		2,288,719	2,297,500	2,118,272	1,724,192
Carbon Dioxide Emitted, STPY	2,353,363	237,382	228,640	407,868	801,947
Cost of Production, $ per Year					
Capital Charge	$ 87,113,705	$ 112,484,409	$ 119,581,526	$ 115,651,286	$ 110,539,513
Coal Cost	$ 29,265,408	$ 32,423,972	$ 32,423,972	$ 32,423,972	$ 32,423,972
Chemicals/Consumables	$ 2,271,894	$ 2,353,786	$ 2,282,874	$ 2,282,874	$ 2,276,798
Maintenance Cost, Materials&Labor	$ 9,218,382	$ 11,903,112	$ 12,654,130	$ 12,238,231	$ 11,697,303
Operating Labor and Administration	$ 8,688,493	$ 8,830,182	$ 8,830,182	$ 8,830,182	$ 8,830,182
Total Cost of Production, $ per year	**$136,557,882**	**$ 167,995,461**	**$175,772,684**	**$ 171,426,546**	**$ 165,767,768**
Net Power Produced, kW_e	424,540	384,029	377,027	396,208	411,710
Cost of Electricity, cents per kWhr	4.59	6.24	6.65	6.17	5.75
% Increase in COE		36%	45%	35%	25%
Avoided Capture Cost, $ per Ton of CO_2	**$ -**	**$ 25**	**$ 32**	**$ 27**	**$ 25**
Avoided Capture Cost, $ per Ton of Carbon	**$ -**	**$ 6.9**	**$ 8.6**	**$ 7.3**	**$ 6.8**
Cost of Shift,$ per Ton of CO2 Captured		$ 6.26	$ 6.05	$ 7.80	$ 7.34

Table 1- 5 Utilities Summary

Item No	Item Name	Load BHP		Elect. Power	Steam, M Pounds per Hour	Cooling Water		Refrigeration	
		Norm.	Max (3).	KW	___ psig	CW, MMbtu/hr	C.W. circ. GPM (2)	NH3, MMbtu/hr	Tons of Refrig.
	Exchangers								
E-100	Syngas Compressor CW Cooler					19.46	2,595		
E-101	Syngas Chiller							17.26	1,438
E-102A	Hydrate Flash Reactor A								
E-102B	Hydrate Flash Reactor B								
E-103	Flash Gas Cooler					3.81	508		
E-104	Recycle Water Chiller							68.0	5,667
E-107A	Acid Gas Compr 1st Intercooler					0.20	20		
E-107B	Acid Gas Compr 2nd Intercooler					0.09	9		
E-107C	Acid Gas Compr 3rd Intercooler					0.08	8		
E-110	CO2 Sequestration Comp Aftercooler					46.89	6,252		
E-201	Ammonia Condenser					94.36	10,318		
H-101	Dryer Regeneration Heater			180					
	Hydrate Formation Reactors								
	Hydrate Formation Reactor							401.50	33,458
	Compressors								
K-100	Flash Gas Compressor	1,573		1,235		0.08	11		
K-101	NH_3 Compressor - Stage 1	23,279		18,280		1.18	158		
K-102	NH_3 Compressor - Stage 2	3,260		2,560		0.17	22		
K-103	Acid Gas Compressor	97		76		0.00	1		
K-104A	Syngas Compressor	14,216		11,163		0.72	96		
K-105	Syngas Expander	(6,305)		(4,468)		0.32	43		
K-107	CO2 Sequestration Compressor	8,324		6,536		0.42	56		
	Pumps								
P-101	Chilled Water Recycle Pumps	7,768		6,439					
P-102	Cooling Water Pumps	411		341					
P-103	Lube Oil Pumps	20		17					
P-104	Demineralized Water Pumps	12.5		10.4					
	TOTAL	52,656		42,370		167.79	20,097	486.76	40,563

NOTES:
1 All Figures shown above represent normal utility usage requirements except:
() indicates normal utility make
* indicates intermittent usage or make, not included in totals
2 Cooling water supply temperature is 80 F. Makeup water to cooling tower is not shown
3 Utility consumption for max. load conditions is not shown.

Table 1- 6 Equipment List

Equipment	No. ea.	Size/Duty	Material Tube/Shell	Temperature (°F) Tube/Shell	Pressure (psig) Tube/Shell
Reactors					
Hydrate Formation Reactors	10	402 MM Btu/hr total	304 SS/CS	200/200	1920/235
Hydrate Flash Reactor A	8	444.5 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/680
Hydrate Flash Reactor B	2	38.3 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/315
Exchangers					
1st Stg Syngas CW Cooler	1	19.5 MM Btu/hr	CS/CS	285/250	1900/75
Syngas Chiller	1	17.3 MM Btu/hr	304 SS/CS	200/200	1,100/235
Recycle Water Chiller	1	68.0. MM Btu/hr	304 SS/CS	200/200	300/235
Flash Gas Cooler	1	3.8 MM Btu/hr	CS/CS	285/200	680/50
Syngas/Product Gas Exch.	1	8.8 MM Btu/hr	CS /304 SS	250/250	1900/1770
CO_2 Sequestration Cooler	1	47.7 MM Btu/hr	CS/CS	375/200	2,400/50
Ammonia Condenser	4	71.6 MM Btu/hr	CS/CS	200/360	75/235
Acid Gas Compressor 1st Stage Intercooler	1	0.20 MM Btu/hr	CS/304 SS (Clad)	200/350	50/50
Acid Gas Compressor 2nd Stage Intercooler	1	0.09 MM Btu/hr	CS/304 SS (Clad)	200/350	50/110
Acid Gas Compressor 3rd Stage Intercooler	1	0.08 MM Btu/hr	CS/304 SS (Clad)	200/350	50/320
Syngas/Expander Preheat Exch	1	21.4 MM Btu/hr	CS/304 SS (Clad)	250/250	1920/1760
Regeneration Heater-Electric	1	170 kW	CS/CS	400	15
Vessels					
Syngas KO Drum	1	8'ID x 10' T-T	304 SS Clad	250	1,055
Slurry/Gas Separator	1	12'ID x 33.5' T-T	304 SS Clad	250	1735
Flash Gas Compr KO Drum	1	4.0'ID x 7.5' T-T	304 SS Clad	120	300
Acid Gas Compr KO Drum	1	1'ID x 5' T-T	304 SS Clad	160	50
Acid Gas 1st Stg Compr KO Drum	1	1'ID x 3' T-T	304 SS Clad	160	100
Acid Gas 2nd Stg Compr KO Drum	1	1'ID x 3' T-T	304 SS Clad	160	230
Acid Gas Product KO Drum	1	1'ID x 3' T-T	304 SS Clad	160	670
Mol Sieve Dryer	2	6.5'ID x 17.5' T-T	304 SS Clad	450	670
Ammonia Separator	1	4'ID x 17' T-T	CS	200	235
Ammonia Surge Drum	1	12'ID x 34' T-T	CS	200	235
Ammonia Compr KO Drum	1	14'ID x41' T-T	CS	200	235
Compressors					
Flash Gas Compressor	1	2,000 HP	CS	350	625
CO2 Sequestration Compressor	1	10,000 HP	CS	350	2400
Acid Gas Compressor	1	150 HP	CS	350	610
NH3 Refrig 1st Stage Compressor	1	27,500 HP	CS	200	136
NH3 Refrig 2nd Stage Compressor	1	4,000 HP	CS	350	245
1st Stg Syngas Compressor	1	16,000 HP	CS	350	1920
1st Stg Syngas Expander	1	5,400 HP	CS	250	1900

Equipment	No. ea.	Size/Duty	Material	Temperature (°F)	Pressure (psig)
Pumps					
Chilled Water Recycle Pumps (3-50% pumps)	3	3700 GPM/ 8,200 HP each	304 SS	200	1100
Cooling Water Pumps (3-50% pumps)	3	12,000 GPM/350 HP each	CS	200	100
Demineralized Water Pumps (2-100% pumps)	2	0.14 GPM/0.03 HP each	CS		
Lube Oil Pumps (3-50% pumps)	3	975 GPM/24 HP each	CS		

Figure 1
CO_2 Recovery vs Capital Cost and Energy Consumption

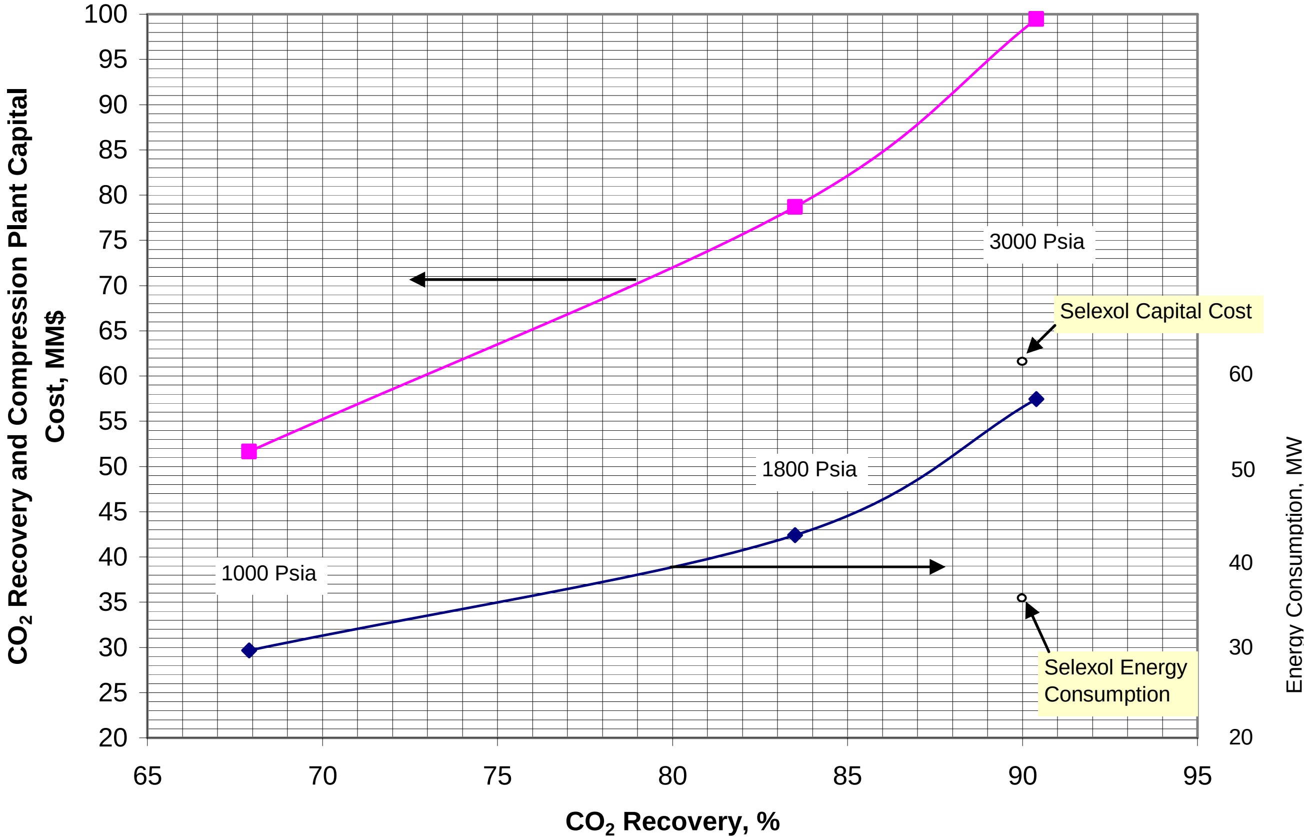


Figure 2
CO_2 Recovery vs Cost of Electricity

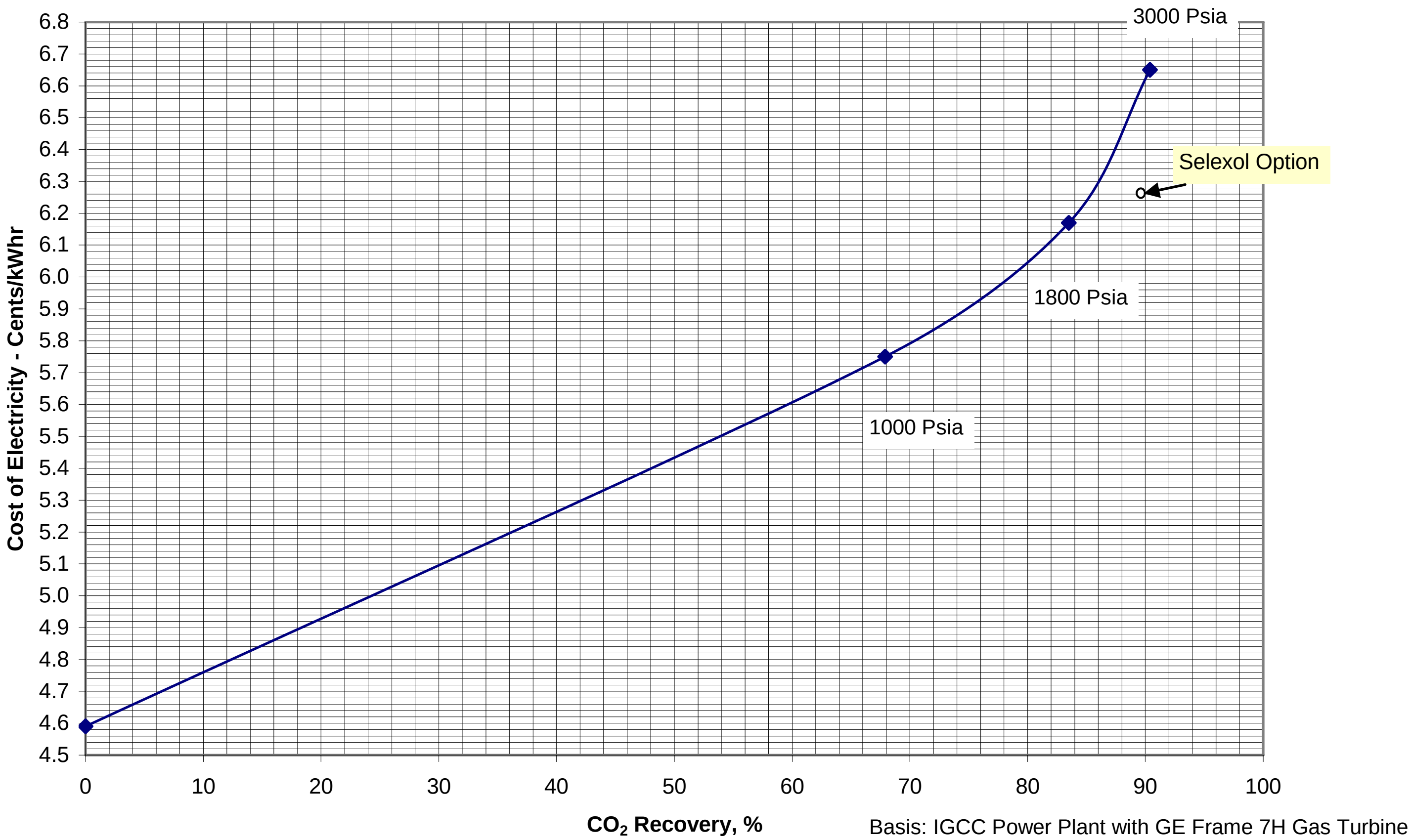


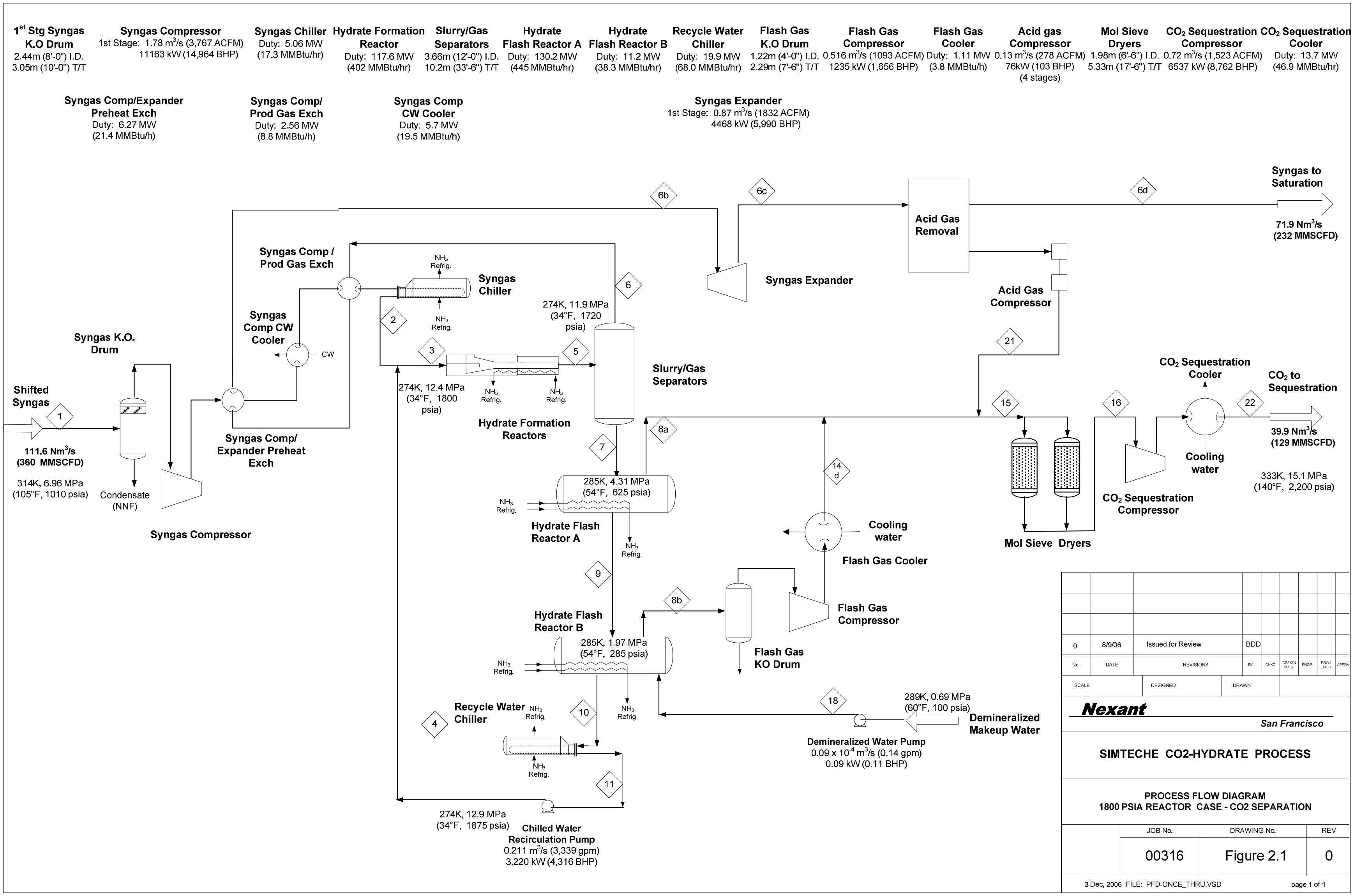
Table 1-7 Material Balance

ONCE THROUGH CASE HEAT & MATERIAL BALANCE - LANL HYDRATE CORRELATION

Coal	**Illinois No. 6**	**FLASH A PRESSURE, PSIA**	**625**		Treated Gas to GT H2S Spec = 40 ppmV
Wt% Sulfur	**2.82 wt% dry(2.51wt% S As Received)**	**FLASH B PRESSURE, PSIA**	**285**	**SR**	**0.835**

Stream Numbers		Strm 1	Strm 1	Strm 1A	Strm 1A	Strm 1B	Strm 1B	Strm 2	Strm 2	Strm 3	Strm 3	Strm 4	Strm 4	Strm 5	Strm 5	Strm 6	Strm 6	Strm 7	Strm 7	Strm 8A	Strm 8A	Strm 8B	Strm 8B	Strm 9	Strm 9	Strm 10	Strm 10
		Sat'd Syngas to KO drum	Sat'd Syngas to KO drum	Sat'd Syngas to chiller	Sat'd Syngas to chiller	KO drum Condensate	KO drum Condensate	Chilled Syngas	Chilled Syngas	Hform Rx Feed	Hform Rx Feed	Nucleated Water	Nucleated Water	Reactor Effluent	Reactor Effluent	Treated Syngas to AGR	Treated Syngas to AGR	Slurry fr HFR-A	Slurry fr HFR-A	CO_2 Vapor fr HFR-A	CO_2 Vapor fr HFR-A	CO_2 Vapor fr HFR-B	CO_2 Vapor fr HFR-B	Water fr fr HFR-A	Water fr fr HFR-A	Water fr fr HFR-B	Water fr fr HFR-B
	Mol Wt	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h
H_2	2.02	21,418.72	43,176	21,418.72	43,176	-	-	21,418.72	43,176	21,418.75	43,176	0.03	0	21,418.75	43,176	21,267.58	42,871	151.16	305	149.33	301	1.81	4	1.84	4	0.03	0
CO_2	44.01	16,381.13	720,930	16,381.13	720,930	-	-	16,381.13	720,930	19,509.89	858,626	3,128.76	137,696	4,180.57	183,986	2,695.83	118,643	1,484.74	65,343	9,866.14	434,207	3,819.16	168,080	6,947.91	305,776	3,128.75	137,696
H_2S	34.08	248.44	8,466	248.44	8,466	-	-	248.44	8,466	463.93	15,809	215.49	7,343	34.02	1,159	13.87	473	20.15	687	145.80	4,968	88.77	3,025	304.25	10,368	215.49	7,343
H_2O Vap	18.02	2.11	38	2.11	38	-	-	2.11	38	2.11	38	-	-	2.11	38	1.39	25	-	-	3.23	58	2.77	50	-	-	-	-
Ar	39.95	292.72	11,694	292.72	11,694	-	-	292.72	11,694	292.72	11,694	0.00	0	292.72	11,694	290.50	11,605	2.22	89	2.19	88	0.02	1	0.03	1	0.00	0
N2	28.01	289.32	8,105	289.32	8,105	-	-	289.32	8,105	289.32	8,105	0.00	0	289.32	8,105	287.13	8,043	2.19	61	2.17	61	0.02	1	0.02	1	0.00	0
CO	28.01	382.31	10,709	382.31	10,709	-	-	382.31	10,709	382.31	10,709	0.00	0	382.31	10,709	377.97	10,587	4.34	122	4.27	120	0.07	2	0.07	2	0.00	0
CH4	16.04	537.79	8,628	537.79	8,628	-	-	537.79	8,628	537.80	8,628	0.01	0	537.80	8,628	528.33	8,476	9.47	152	9.24	148	0.23	4	0.23	4	0.01	0
COS	60.07	0.04	2	0.04	2	-	-	0.04	2	0.05	3	0.01	1	0.05	3	0.02	1	0.03	2	0.01	1	0.01	0	0.02	1	0.01	1
NH3	17.03	3.40	58	3.40	58	-	-	3.40	58	324.09	5,519	320.69	5,461	324.09	5,519	1.58	27	322.51	5,492	0.98	17	0.84	14	321.53	5,476	320.69	5,461
H_2O Liq	18.02	-	-	-	-	-	-	-	-	183,100.02	3,298,583	183,100.02	3,298,583	88,544.63	1,595,149	-	-	88,545.35	1,595,162	-	-	-	-	183,097.5	3,298,538	183,100.0	3,298,583
CO_2 Hydrate	152.10	-	-	-	-	-	-	-	-	-	-	-	-	15,329.32	2,331,605	-	-	15,329.32	2,331,605	-	-	-	-	-	-	-	-
H_2S Hydrate	142.17	-	-	-	-	-	-	-	-	-	-	-	-	429.91	61,119	-	-	429.91	61,119	-	-	-	-	-	-	-	-
Total		39,555.98	811,805	39,555.98	811,805	-	-	39,555.98	811,805	226,320.98	4,260,890	186,765.00	3,449,085	131,765.60	4,260,890	25,464.20	200,751	106,301.4	4,060,139	10,183.36	439,968	3,913.70	171,181	190,673.4	3,620,170	186,765.0	3,449,085
Temperature, F		105		105		105		34		34		34		34		34		34		54		54		54		54	
Pressure, psia		1,010		1,010		1,010		1,800		1,800		1,850		1,720		1,720		1,720		625		285		625		285	

Stream Numbers		Strm 11	Strm 11	Strm 14A	Strm 14A	Strm 14B	Strm 14B	Strm 14C	Strm 14C	Strm 14D	Strm 14D	Strm 15	Strm 15	Strm 16	Strm 16	Strm 17	Strm 17	Strm 18	Strm 18	Strm 6A	Strm 6A	Strm 21	Strm 21	Strm 22	Strm 22
		Chilled Recycle Water	Chilled Recycle Water	FlashGas Compr KO Liq	FlashGas Compr KO Liq	FlashGas Compr Feed	FlashGas Compr Feed	FlashGas Compr Discharge	FlashGas Compr Discharge	Cooled FG Compr Discharge	Cooled FG Compr Discharge	Total CO_2 to Dryer	Total CO_2 to Dryer	CO2 Seq Compr Suction	CO2 Seq Compr Suction	Total Condensate	Total Condensate	Demin Water Makeup	Demin Water Makeup	Treated Gas to GT	Treated Gas to GT	Acid Gas fr AGR	Acid Gas fr AGR	Total CO2 to Seques	Total CO2 to Seques
	Mol Wt	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h	lbmol/h	lb/h
H_2	2.02	0.03	0	-	-	1.81	4	1.81	4	1.81	4	151.14	305	151.14	305	-	-	-	-	21,267.58	42,871	-	-	151.14	305
CO_2	44.01	3,128.75	137,696	-	-	3,819.16	168,080	3,819.16	168,080	3,819.16	168,080	13,736.25	604,530	13,736.25	604,530	-	-	-	-	2,644.88	116,401	50.95	2,242	13,736.25	604,530
H_2S	34.08	215.49	7,343	-	-	88.77	3,025	88.77	3,025	88.77	3,025	247.42	8,431	247.42	8,431	-	-	-	-	1.02	35	12.85	438	247.42	8,431
H_2O Vap	18.02	-	-	-	-	2.77	50	2.77	50	2.77	50	17.89	322	17.89	322	-	-	-	-	33.56	605	11.89	214	17.89	322
Ar	39.95	0.00	0	-	-	0.02	1	0.02	1	0.02	1	2.22	89	2.22	89	-	-	-	-	290.50	11,605	-	-	2.22	89
N2	28.01	0.00	0	-	-	0.02	1	0.02	1	0.02	1	2.19	61	2.19	61	-	-	-	-	287.13	8,043	-	-	2.19	61
CO	28.01	0.00	0	-	-	0.07	2	0.07	2	0.07	2	4.34	122	4.34	122	-	-	-	-	377.97	10,587	-	-	4.34	122
CH4	16.04	0.01	0	-	-	0.23	4	0.23	4	0.23	4	9.46	152	9.46	152	-	-	-	-	528.33	8,476	-	-	9.46	152
COS	60.07	0.01	1	-	-	0.01	0	0.01	0	0.01	0	0.02	1	0.02	1	-	-	-	-	0.02	1	-	-	0.02	1
NH3	17.03	320.69	5,461	-	-	0.84	14	0.84	14	0.84	14	1.82	31	1.82	31	-	-	-	-	1.58	27	-	-	1.82	31
H_2O Liq	18.02	183,100.02	3,298,583	-	-	-	-	-	-	-	-	-	-	-	-	-	-	5.28	95	-	-	-	-	-	-
CO_2 Hydrate	152.10	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H_2S Hydrate	142.17	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total		186,765.00	3,449,085	-	-	3,913.70	171,181	3,913.70	171,181	3,913.70	171,181	14,172.76	614,043	14,172.76	614,043	-	-	5.28	95	25,432.57	198,651	75.69	2,894	14,172.76	614,043
Temperature, F		34		54		54		182		110		68		68		100		54		105		244		140	
Pressure, psia		280		285		285		630		625		625		615		625		295		922		625		2,200	


1st Stg Syngas K.O Drum
2.44m (8'-0") I.D.
3.05m (10'-0") T/T
Syngas Compressor
1st Stage: 1.78 m3/s (3,767 ACFM)
11163 kW (14,964 BHP)
Syngas Chiller
Duty: 5.06 MW
(17.3 MMBtu/hr)
Hydrate Formation Reactor
Duty: 117.6 MW
(402 MMBtu/hr)
Slurry/Gas Separators
3.66m (12'-0") I.D.
10.2m (33'-6") T/T
Hydrate Flash Reactor A
Duty: 130.2 MW
(445 MMBtu/hr)
Hydrate Flash Reactor B
Duty: 11.2 MW
(38.3 MMBtu/hr)
Recycle Water Chiller
Duty: 19.9 MW
(68.0 MMBtu/hr)
Flash Gas K.O Drum
1.22m (4'-0") I.D.
2.29m (7'-6") T/T
Flash Gas Compressor
0.516 m3/s (1093 ACFM)
1235 kW (1,656 BHP)
Flash Gas Cooler
Duty: 1.11 MW
(3.8 MMBtu/h)
Acid gas Compressor
0.13 m3/s (278 ACFM)
76kW (103 BHP)
(4 stages)
Mol Sieve Dryers
1.98m (6'-6") I.D.
5.33m (17'-6") T/T
CO2 Sequestration Compressor
0.72 m3/s (1,523 ACFM)
6537 kW (8,762 BHP)
CO2 Sequestration Cooler
Duty: 13.7 MW
(46.9 MMBtu/hr)
Syngas Comp/Expander Preheat Exch
Duty: 6.27 MW
(21.4 MMBtu/h)
Syngas Comp/ Prod Gas Exch
Duty: 2.56 MW
(8.8 MMBtu/h)
Syngas Comp CW Cooler
Duty: 5.7 MW
(19.5 MMBtu/h)
Syngas Expander
1st Stage: 0.87 m3/s (1832 ACFM)
4468 kW (5,990 BHP)
Syngas to Saturation
71.9 Nm3/s
(232 MMSCFD)
6b
6c
6d
Acid Gas Removal
Syngas Comp / Prod Gas Exch
NH3 Refrig.
Syngas Chiller
6
274K, 11.9 MPa
(34°F, 1720 psia)
Syngas Expander
Acid Gas Compressor
Syngas Comp CW Cooler
CW
2
3
5
21
Syngas K.O. Drum
Slurry/Gas Separators
CO2 Sequestration Cooler
CO2 to Sequestration
Shifted Syngas
274K, 12.4 MPa
(34°F, 1800 psia)
Hydrate Formation Reactors
15
16
22
1
8a
111.6 Nm3/s
(360 MMSCFD)
Syngas Comp/ Expander Preheat Exch
7
39.9 Nm3/s
(129 MMSCFD)
Cooling water
314K, 6.96 MPa
(105°F, 1010 psia)
Condensate (NNF)
14d
285K, 4.31 MPa
(54°F, 625 psia)
333K, 15.1 MPa
(140°F, 2,200 psia)
CO2 Sequestration Compressor
Syngas Compressor
Hydrate Flash Reactor A
Cooling water
Mol Sieve Dryers
Flash Gas Cooler
9
8b
Flash Gas Compressor
Hydrate Flash Reactor B
285K, 1.97 MPa
(54°F, 285 psia)
Flash Gas KO Drum
0
8/9/06
Issued for Review
BDD
No.
DATE
REVISIONS
BY
CHKD
DESIGN SUPV
ENGR
PROJ ENGR
APPRV
SCALE:
DESIGNED:
DRAWN:
18
289K, 0.69 MPa
(60°F, 100 psia)
Recycle Water Chiller
10
4
Demineralized Makeup Water
Nexant
San Francisco
Demineralized Water Pump
0.09 x 10-4 m3/s (0.14 gpm)
0.09 kW (0.11 BHP)
SIMTECHE CO2-HYDRATE PROCESS
11
PROCESS FLOW DIAGRAM
1800 PSIA REACTOR CASE - CO2 SEPARATION
274K, 12.9 MPa
(34°F, 1875 psia)
Chilled Water Recirculation Pump
0.211 m3/s (3,339 gpm)
3,220 kW (4,316 BHP)
JOB No.
DRAWING No.
REV
00316
Figure 2.1
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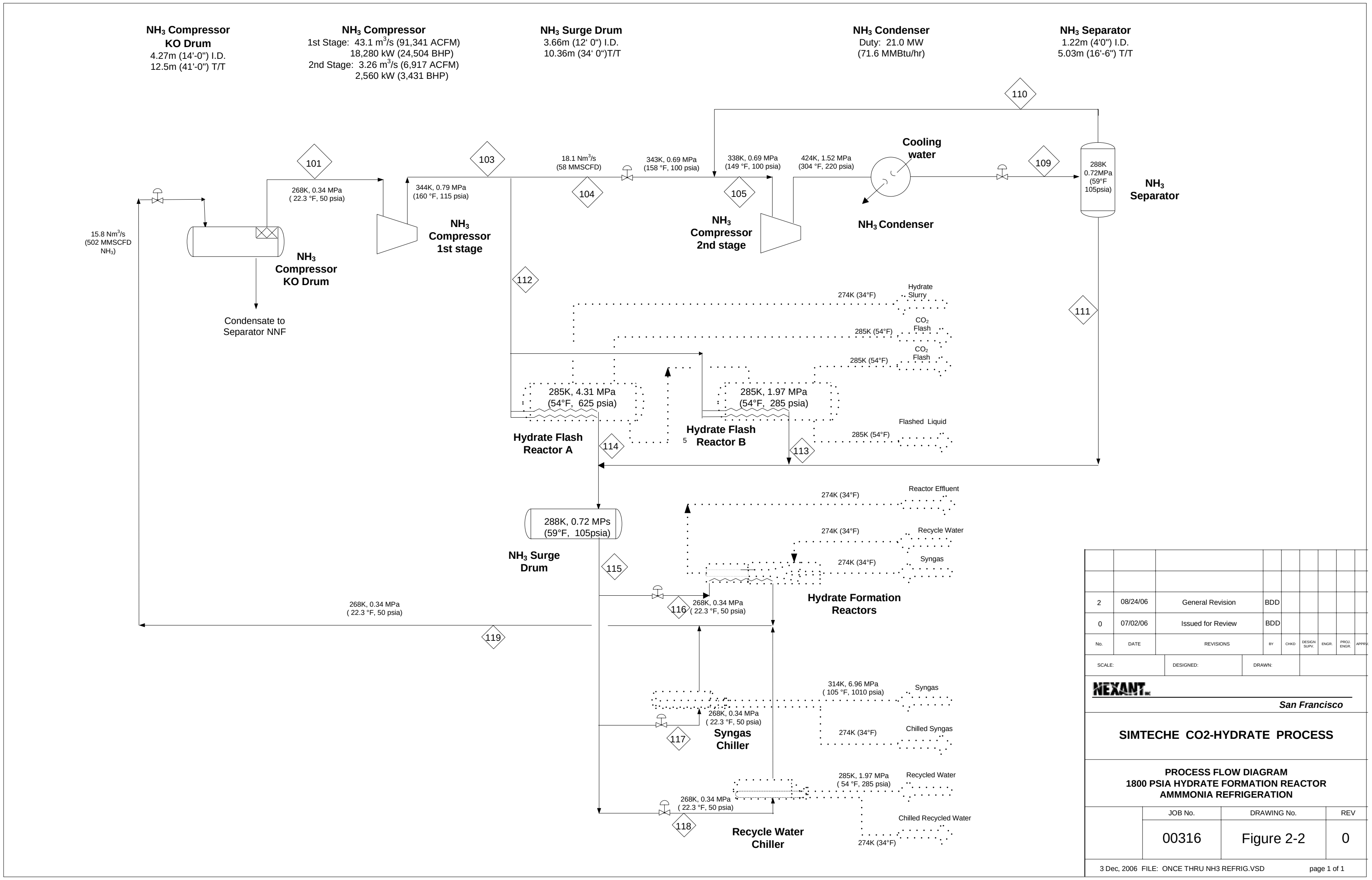
3 Dec, 2006 FILE: PFD-ONCE_THRU.VSD
page 1 of 1


NH3 Compressor KO Drum
4.27m (14'-0") I.D.
12.5m (41'-0") T/T
NH3 Compressor
1st Stage: 43.1 m3/s (91,341 ACFM)
18,280 kW (24,504 BHP)
2nd Stage: 3.26 m3/s (6,917 ACFM)
2,560 kW (3,431 BHP)
NH3 Surge Drum
3.66m (12' 0") I.D.
10.36m (34' 0")T/T
NH3 Condenser
Duty: 21.0 MW
(71.6 MMBtu/hr)
NH3 Separator
1.22m (4'0") I.D.
5.03m (16'-6") T/T
15.8 Nm3/s
(502 MMSCFD NH3)
268K, 0.34 MPa
(22.3 °F, 50 psia)
344K, 0.79 MPa
(160 °F, 115 psia)
18.1 Nm3/s
(58 MMSCFD)
343K, 0.69 MPa
(158 °F, 100 psia)
338K, 0.69 MPa
(149 °F, 100 psia)
424K, 1.52 MPa
(304 °F, 220 psia)
Cooling water
288K
0.72MPa
(59°F
105psia)
NH3 Separator
NH3 Compressor 1st stage
NH3 Compressor 2nd stage
NH3 Condenser
NH3 Compressor KO Drum
Condensate to Separator NNF
Hydrate Slurry
274K (34°F)
CO2 Flash
285K (54°F)
285K, 4.31 MPa
(54°F, 625 psia)
285K, 1.97 MPa
(54°F, 285 psia)
Hydrate Flash Reactor A
Hydrate Flash Reactor B
Flashed Liquid
Reactor Effluent
Recycle Water
Syngas
288K, 0.72 MPs
(59°F, 105psia)
NH3 Surge Drum
Hydrate Formation Reactors
268K, 0.34 MPa
(22.3 °F, 50 psia)
314K, 6.96 MPa
(105 °F, 1010 psia)
Syngas Chiller
Chilled Syngas
285K, 1.97 MPa
(54 °F, 285 psia)
Recycled Water
Chilled Recycled Water
Recycle Water Chiller
101
103
104
105
109
110
111
112
113
114
115
116
117
118
119
2 08/24/06 General Revision BDD
0 07/02/06 Issued for Review BDD
No. DATE REVISIONS BY CHKD DESIGN SUPV. ENGR. PROJ. ENGR. APPRV.
SCALE: DESIGNED: DRAWN:
NEXANT
San Francisco
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
1800 PSIA HYDRATE FORMATION REACTOR
AMMMONIA REFRIGERATION
JOB No. DRAWING No. REV
00316 Figure 2-2 0
3 Dec, 2006 FILE: ONCE THRU NH3 REFRIG.VSD page 1 of 1

Task 6 – Two-Step 2400 Psia Hydrate Formation Reactor Operating Pressure

Objective

The results of Task 4 (3000 psia hydrate formation reactor operating pressure) indicates that considerably higher capital and operating costs will be incurred if the Simteche process is to recover 90% of the CO_2 in a single stage hydrate formation reactor. A way to reduce this cost is to utilize a scheme with a two-step hydrate formation reactor without any separation equipment between them but operating at different temperatures. In this way, a first step of bulk hydration formation can be achieved at a higher temperature followed by the final removal in a lower temperature reactor. This will reduce the refrigeration loads and as a result reduce the capital and operating costs. Experimental observations at LANL have shown that hydrates form at considerably higher temperatures when the reactors are operated at higher CO_2 partial pressures.

This 'Two-Step Hydrate Formation' is the latest concept developed by Simteche. Preliminary back-of-the-envelope calculations by Simteche showed that 68% of the Hydrates can be recovered at 7^oC and the other 22% can be captured at 0.5^oC (for a total of 90%) if the Hydrate reactor is operated at around 2400 psia. This higher operating pressure can be achieved by operating the gasifier at as high a pressure as possible, supplemented by additional syngas compression.

The objective of this task is to revise the Simteche CO_2 Hydrate process scheme allowing the first section of the hydrate reactor to be operated at 7^oC and 2400 psia, followed by a second reactor section at 0.5 ^oC and 2360 psia; then carry out design heat, material and utility balance calculations, and cost estimation to allow the technical and economic analyses of this new process scheme. The results are to be compared to other process schemes as well as the Selexol process investigated in the Engineering Analysis Study, February 2006.

DOE/NETL accepted a Nexant/Simteche request to proceed with the subject study, which indicated that it would yield improved results compared to the replaced original Contract Mod 017 Added Task #6 work scope. The original task had an objective of studying an 'Alternative Heat Removal' scheme for the Simteche process.

Summary

The capital cost and energy consumption of this 'Two-Step Hydrate Formation' scheme is the lowest of all the Simteche schemes studied thus far that can remove 90% CO_2 in a single stage. This is shown in Table 1.4 where the yearly cost of production, including capital charge and operating costs, for the various schemes are presented. The current 'Two-Step Hydrate Formation' scheme has the lowest cost of electricity of 6.36 cents/kWhr. It is competitive with Selexol, which also has a 36% increase in the cost of electricity compared to the base case without CO_2 capture. The 'Two-Step Hydrate Formation' scheme, allowing the process to be operated at a higher pressure, in sequential temperature steps, results in lower refrigeration operating and capital costs. The subject techno-economic analysis comparing all the Simteche schemes that recover 90% CO_2 were performed on a fully integrated IGCC (EPRI prototype) plant design.

Table 1.1 summarizes all the major design and cost details for the various cases. Relative economic comparisons were performed on a cost of electricity basis.

Two-Step Hydrate Formation Reactor Case – Process Scheme

This scheme is similar to the 1800 Psia once-thru scheme except there are two reactors in series each operating at a different temperature. In this scheme the syngas is compressed to 2450 psia in a single stage compressor. This is followed with cooling by exchange with syngas product, cooling water, and ammonia to 45^oF (see Fig 2.1). It is then mixed with recycled chilled water and sent to two Hydrate Formation reactors in series. Here the mixture is chilled by exchange with ammonia and the resultant CO_2 and H_2S hydrates are formed. The first reactor is operated at a 7^oC ($45\ ^oF$) temperature where approximately two-thirds of the CO_2 is converted to hydrate. The reactor effluent without any intermediate separation is the fed to the second reactor which operates at 0.5^oC (32.9^oF). The first reactor where the bulk of the heat of formation is removed is chilled with an ammonia stream at a higher pressure of 65 psia compared to the base case of 50 psia and thus requires 23% less refrigeration horsepower.

The effluent from the second reactor is sent to a slurry/gas separator which operates at 2320 psia and 32.9^oF and where the phases are separated. The gas stream is heated first by exchange with syngas feed and then by exchange with the first stage Syngas compressor discharge to preheat it to $266\ ^oF$ before being sent to single stage turbo expander. This preheat is sufficient to avoid any ice or other solids formation in the expander and downstream equipment and provides additional power. The effluent from the expander exhausts at 922 psia and $127\ ^oF$, where it is sent to an acid gas removal unit which removes H_2S to meet the syngas product spec of 40 ppm sulfur. The final product is then sent to syngas humidification. A small waste gas stream from the acid gas removal unit containing H_2S and a small amount of CO2 is compressed to 630 psia.

The liquid slurry from the slurry/gas separator is letdown in pressure to 625 psia and the slurry is heated to $54\ ^oF$ in Hydrate Flash Reactor A. Here the Hydrate is disassociated to CO_2, H_2S and water and 60% of the CO_2 is flashed off and sent to the Mol Sieve Gas Dehydration unit. The heating is accomplished by condensing part of the 1st stage NH3 refrigeration compressor effluent. The remaining liquid containing water, CO_2 and H_2S is letdown in pressure to 285 psia and sent to Hydrate Flash Reactor B. Here the mixture is heated again and additional CO_2 and H_2S are flashed off. This heating is also accomplished by condensing part of the 1st stage NH3 refrigeration compressor effluent. This off gas is compressed in a Flash Gas compressor to 630 psia and mixed with the off gas from Hydrate Flash Reactor A and the waste gas from the acid gas removal unit. This mixture is then sent to a Mol Sieve gas Dehydration unit to meet the product CO_2

dryness specification. The dried CO_2 is then compressed in the CO_2 sequestration compressor and cooled and sent offsite for sequestration.

The liquid from the Hydrate Flash Reactor B containing mostly water and dissolved CO_2 and H2S is then chilled to 45 ^oF and pumped back to the inlet of the Hydrate Formation reactor.

Techno-economic Analysis - Methodology

An existing spreadsheet model, utilized in designing the cases included in the Engineering Analysis Report of February 2006 was used to develop the material balances. It incorporated the Henry's law coefficients to handle the flashes with the hydrate present. An additional simulation model was built using Hysys to develop the heat and mass balance and facilitate sizing and costing the plant. The Hysys model was modified to handle the equilibrium flashes using the same Henry's law constants that were employed in the spreadsheet. The simulation model was unable to recognize hydrate formation or hydrate disassociation. The heat balance was adjusted manually with the heats of hydrate formation as determined below. The heat balance around the hydrate formation reactor and the hydrate heater were handled separately using Hysys heat capacities and heats of formation of the hydrates. The total CO_2 heats of hydrate formation (vapor phase to solid hydrate) was assumed to be 62 kJ/gmol[1]. Heat of hydrate formation of CO_2 and H_2S (solid hydrate to vapor phase) were assumed to be the same at 2400 psia as at 1000 psia. The H_2S heats of hydrate formation were assumed to be the same as CO_2 on a molar basis since they had similar hydration numbers and any deviation was expected to be small since the major contributor is the heat of fusion of water. In addition the heat of hydration of H_2S represents only 2% of the hydrate formation reactor duty. These heats of hydrate formation will need to be verified by experimental data if this case is to be pursued in the future.

The property packages used in Hysys are known to be accurate in the 1000 to 2500 psia conditions of the syngas compression and expansion.

Because the high pressure flash takes place at 2320 psia well above the syngas product delivery pressure of 900 psia, additional energy can be recovered from this stream by means of a gas turbo expander to reduce the syngas pressure to 920 psia. However expanders can result in low discharge temperatures where ice and hydrates can form. To avoid the potential for low temperatures the expander feed gas is preheated with exchange with the syngas compressor discharge streams. This has the following advantages:

- Reduces the CW requirement for the syngas compressors
- Increases the power attainable from the gas turbines
- Eliminates hydrate formation and localized freezing in the downstream Acid Gas Removal unit

[1] "Syngas Upgrading – A Low Temperature Approach, Phase 1 Report, June 2002" by Los Alamos National Laboratory, Simteche, and Nexant

The 2400 psia 2 step case along with its comparison with the other cases are presented in the following:

Table 1-1 - Economic Comparative Analysis
Table 1-2 - Equipment Cost Estimates
Table 1-3 - Power Production Summary
Table 1-4 - Carbon Control Cost
Table 1.5 - Utility summary
Table 1.6 - Equipment list
Table 1.7 - Material Balance
Figure 2.1 - Simplified PFD CO_2 Recovery Section
Figure 2.2 - Simplified PFD NH_3 Refrigeration Section

Results

As can be seen in Table 1.1, the 2400 psia 2-step reactor scheme has the lowest refrigeration load of all the Simteche cases. This is because most of the cooling is done at a higher refrigeration temperature than the other once-thru Simteche cases. However, it requires an additional net 10MW of power for the syngas compressors after taking credit for the syngas expander. Table 1.1 also shows that this case has the lowest cost of electricity at 6.36 cents/kWhr of the three single stage Simteche cases. Table 1.2 shows that this case has the lowest installed cost for the CO_2 removal and compression plants of the three cases.

This case also has the lowest yearly production costs of the three single stage Simteche cases as summarized in Table 1.4. In addition, it has the highest electric power generating fuel efficiency (35.5%) of the three cases and the lowest percentage of parasitic load consumption (9.5%) as can be seen in Table 1.3.

Table 1.4 compares the avoided capture cost per ton of CO_2 for all the Simteche cases as well as the Selexol option. As can be seen, the avoided capture cost of CO_2 is $27/ ton for this case as opposed to $31 to 32 $/ ton for the other single stage cases. This is, however, greater than the $24/ ton cost for the two stage promoter case which used an optimistic hydration number and greater than the $25/ ton cost for the Selexol case. Thus this case is slightly less competitive with the Selexol option.

It is interesting to note that the lowest avoided CO_2 cost is $25/ ton for the Simteche once-thru 1000 psia base case. This shows that the Simteche technology works best at bulk CO_2 removal but becomes less competitive as recovery rates approach 90%.

Also to be noted in Tables 1.1 through 1.4, the two stage promoter case has lower capital and operating costs, higher payout and a lower cost of CO_2 capture than all the one stage Simteche processes. This is based on early preliminary assumptions for the hydration number of the promoter. During the course of these studies, it was noted that this hydration number may be optimistic based on the recent data from LANL. It could be much higher which would increase the reactor cooling and refrigeration duty, thereby increasing operating and capital costs and altering the conclusions for this case.

Table 1-1 Economic Comparative Analysis

Case		3000 Psia Hydrate Form Reactor	2400 Psia 2 Step Hydrate Form Reactor	1-Stage Promoter	Selexol
CO2 Recovery		**90**	**90**	**90**	**90**
Capital Cost (Note 1)	MM$	99.47	84.91	100.13	61.92
Power Revenues					
Gas Turbine Power	KW	341577	341529	345355	333720
Steam Turbine Power (Note 1)	KW	145814	145474	150362	139294
Generator Loss	KW	(7250)	(7249)	(7330)	(7330)
Turboset Power	KW	480141	479754	488387	465684
Fuel Gas Expander Power	KW	9162	9115	9005	9005
Total Gross Power Generated	KW	489303	488869	497392	474689
Auxiliary Loads	KW				
1st Stg NH3 Compressors	KW	19672	1122	38203	3787
2nd Stg NH3 Compressors	KW	2869	13572	6007	
3rd Stg NH3 Compressors	KW		2646		
Acid Gas Compressors	KW	46	62	90	
Flash Gas Compressors	KW	1308	1393	1009	9726
1st stg Syngas Compressor	KW	10716	17303		
2nd stg Syngas Compressor	KW	10443	0		
CO2 Sequestration Compressor	KW	7044	7113	7696	15380
1st Stg Product Gas Expander	KW	(4026)	(6982)		
2nd Stg Product Gas Expander	KW	(3536)			
Solvent Pumps	KW				5042
Cooling Water Pumps	KW	459	391	623	501
Recycle Water Pump	KW	12192	9635	6282	
Dryer Regen Heater	KW	190	182	182	1330
Other Loads	KW	22	17	17	
MDEA AGR		47	47	47	
CO2 Removal Plant Loads	KW	57446	46500	60156	35766
Cooling Tower Fans	KW	1150	1130	1198	1162
Balance of Plant	KW	53680	53680	53680	53680
Total Auxiliary loads	KW	112276	101311	115034	90608
OnStream Factor		0.8	0.8	0.8	0.8
Net Power	KW	377027	387558	382358	384081
Net Cost of Electricity	cents/kWhr	6.65	6.36	6.56	6.24

Note 1 Cooling tower capital cost not included

Table 1-2 Equipment Cost Estimates

CO_2 Removal and Compression	Selexol	SIMTECHE			
		2-Stage Promoter	*1-Stage Promoter*	*1-Step 3000 Psia Reactor*	*2-Step 2400 Psia Reactor*
CO_2 Separation Ratio	90%	90%	90%	90%	90%
Total Installed Cost, $ 2Q 2005					
Vessels, Exchangers, Pumps	$ 39,987,946	$ 48,049,292	$ 65,113,854	$ 64,933,505	$ 56,453,167
Refrigeration Compression	$ 3,046,633	$ 19,773,269	$ 28,098,849	$ 16,174,463	$ 11,547,918
CO_2 + Other Compression	$ 17,817,452	$ 8,323,126	$ 6,291,840	$ 17,749,899	$ 16,294,846
MDEA AGR	Not Required	$ 555,000	$ 555,000	$ 555,000	$ 555,000
Column Packing, Initial Fill	$ 1,069,338	$ 67,874	$ 71,762	$ 59,385	$ 59,385
Total CO2 Removal and Compression	**$61,921,368**	**$76,768,561**	**$100,131,304**	**$99,472,251**	**$84,910,316**

Table 1-3 Power Production Summary

Power Production Summary	No Capture	Selexol	SIMTECHE			
			2-Stage Promoter	*1-Stage Promoter*	*1-Step 3000 Psia Reactor*	*2-Step 2400 Psia Reactor*
CO_2 Separation Ratio	0%	90%	90%	90%	91%	90%
Gross Plant Power, kW_e	474,040	474,689	497,392	497,392	489,303	488,854
Auxilliary Power Loads						
CO_2 Capture and Compression		35,766	44,265	59,928	57,446	46,500
Power Plant	49,500	54,894	54,878	54,878	54,830	54,810
Net Plant Power, kW_e	424,540	384,029	398,249	382,586	377,027	387,544
Net Efficiency, %HHV	43.0	35.2	36.5	35.1	34.6	35.5
Heat Rate, BTU/kWhr	7,936	9,693	9,347	9,729	9,873	9,605
Capture Parasitic Load, %		7.5%	8.9%	12.0%	11.7%	9.5%

Table 1-4 Carbon Control Cost

Carbon Control Costs	No Capture	Selexol	SIMTECHE			
			2-Stage Promoter	*1-Stage Promoter*	*1-Step 3000 Psia Reactor*	*2-Step 2400 Psia Reactor*
CO_2 Separation Ratio	0%	90%	90%	90%	91%	90%
Onstream Factor	80%	80%	80%	80%	80%	80%
Coal Feed (as received), Short Tons per Year	1,009,152	1,118,068	1,118,068	1,118,068	1,118,068	1,118,068
Carbon Dioxide Captured, STPY		2,288,719	2,278,145	2,278,499	2,297,500	2,273,528
Carbon Dioxide Emitted, STPY	2,353,363	237,382	247,995	252,624	228,640	252,612
Cost of Production, $ per Year						
Capital Charge	$ 87,113,705	$ 112,484,409	$ 115,290,529	$119,706,087	$119,581,526	$116,829,320
Coal Cost	$ 29,265,408	$ 32,423,972	$ 32,423,972	$ 32,423,972	$ 32,423,972	$ 32,423,972
Chemicals/Consumables	$ 2,271,894	$ 2,353,786	$ 2,285,469	$ 2,285,469	$ 2,282,874	$ 2,282,874
Maintenance Cost, Materials&Labor	$ 9,218,382	$ 11,903,112	$ 12,200,056	$ 12,667,311	$ 12,654,130	$ 12,362,891
Operating Labor and Administration	$ 8,688,493	$ 8,830,182	$ 8,830,182	$ 8,830,182	$ 8,830,182	$ 8,830,182
Total Cost of Production, $ per year	**$136,557,882**	**$ 167,995,461**	**$ 171,030,207**	**$175,913,021**	**$175,772,684**	**$172,729,239**
Net Power Produced, kW_e	424,540	384,029	398,249	382,586	377,027	387,544
Cost of Electricity, cents per kWhr	4.59	6.24	6.13	6.56	6.65	6.36
% Increase in Cost of Electricity	0	36%	34%	43%	45%	39%
Avoided Capture Cost, $ per Ton of CO2	**$ -**	**$ 25**	**$ 24**	**$ 31**	**$ 32**	**$ 27**
Avoided Capture Cost, $ per Ton of Carbon	**$ -**	**$ 6.9**	**$ 6.4**	**$ 8.3**	**$ 8.6**	**$ 7.5**
Cost of Shift,$ per Ton of Carbon		$ 6.26	$ 5.93	$ 5.97	$ 6.05	$ 6.09

Table 1-5 Utilities Summary

Item No	Item Name	Load BHP		Elect. Power	Cooling Water		Refrigeration	
		Norm.	Max (3).	KW	CW, MMbtu/hr	C.W. circ. GPM (2)	NH3, MMbtu/hr	Tons of Refrig.
	Exchangers							
E-100	Syngas Compressor CW Cooler				30.90	4,120		
E-101	Syngas Chiller						14.56	1,213
E-102A	Hydrate Flash Reactor A							
E-102B	Hydrate Flash Reactor B							
E-103	Flash Gas Cooler				4.29	572		
E-104	Recycle Water Chiller						36.4	3,036
E-107A	Acid Gas Compr 1st Intercooler				0.15	15		
E-107B	Acid Gas Compr 2nd Intercooler				0.07	7		
E-107C	Acid Gas Compr 3rd Intercooler				0.06	6		
E-110	CO2 Sequestration Comp Aftercooler				49.78	6,637		
E-201	Ammonia Condenser				75.32	8,416		
H-101	Dryer Regeneration Heater			182				
	Hydrate Formation Reactors							
	Hydrate Formation Reactor A						314.06	26,172
	Hydrate Formation Reactor B						169.16	14,097
	Compressors							
K-100	Flash Gas Compressor	1,774		1,393	0.09	12		
K-101	NH_3 Compressor - Stage 1	1,429		1,122	0.07	10		
K-102	NH_3 Compressor - Stage 2	17,283		13,572	0.88	117		
K-103	NH_3 Compressor - Stage 2	3,369		2,646	0.17	23		
K-103	Acid Gas Compressor	75		62	0.00	1		
K-104A	Syngas Compressor	22,035		17,303	1.12	150		
K-105	Syngas Expander	(9,853)		(6,982)	0.50	67		
K-107	CO2 Sequestration Compressor	9,058		7,113	0.46	61		
	Pumps							
P-101	Chilled Water Recycle Pumps	11,624		9,635				
P-102	Cooling Water Pumps	472		391				
P-103	Lube Oil Pumps	20		17				
P-104	Demineralized Water Pumps	0.1		0.1				
	MDEA Acid Gas Removal			47				
	TOTAL	57,287		46,500	163.87	20,214	534.21	44,518

ES:

1 All Figures shown above represent normal utility usage requirements except:
() indicates normal utility make
* indicates intermittent usage or make, not included in totals
2 Cooling water supply temperature is 80 F. Makeup water to cooling tower is not shown
3 Utility consumption for max. load conditions is not shown.

Table 1-6 Equipment List

Equipment	No. ea.	Size/Duty	Material Tube/Shell	Temperature (°F) Tube/Shell	Pressure (psig) Tube/Shell
Reactors					
Hydrate Formation Reactors	10	314 MM Btu/hr total	304 SS/CS	200/200	2550/235
Hydrate Formation Reactors	10	169 MM Btu/hr total	304 SS/CS	200/200	2550/235
Hydrate Flash Reactor A	8	474.4 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/680
Hydrate Flash Reactor B	2	40.5 MM Btu/hr	304 SS/304 SS (Clad)	250/200	235/315
Exchangers					
1st Stg Syngas CW Cooler	1	30.9 MM Btu/hr	CS/CS	285/250	2550/75
Syngas Chiller	1	14.6 MM Btu/hr	304 SS/CS	200/200	1,100/235
Recycle Water Chiller	1	36.4 MM Btu/hr	304 SS/CS	200/200	300/235
Flash Gas Cooler	1	4.3 MM Btu/hr	CS/CS	285/200	680/50
Syngas/Product Gas Exch.	1	8.4 MM Btu/hr	CS /304 SS	250/250	2550/2420
CO_2 Sequestration Cooler	1	49.8 MM Btu/hr	CS/CS	375/200	2,400/50
Ammonia Condenser	4	75.3 MM Btu/hr	CS/CS	200/360	75/235
Acid Gas Compressor 1st Stage Intercooler	1	0.15 MM Btu/hr	CS/304 SS (Clad)	200/350	50/50
Acid Gas Compressor 2nd Stage Intercooler	1	0.07 MM Btu/hr	CS/304 SS (Clad)	200/350	50/110
Acid Gas Compressor 3rd Stage Intercooler	1	0.06 MM Btu/hr	CS/304 SS (Clad)	200/350	50/320
Syngas/Expander Preheat Exch	1	32.9 MM Btu/hr	CS/304 SS (Clad)	250/250	2550/2400
Regeneration Heater-Electric	1	182 kW	CS/CS	400	15
Vessels					
Syngas KO Drum	1	8'ID x 10' T-T	304 SS Clad	250	1,055
Slurry/Gas Separator	1	11'ID x 40' T-T	304 SS Clad	250	2355
Flash Gas Compr KO Drum	1	4.5'ID x 7.5' T-T	304 SS Clad	120	300
Acid Gas Compr KO Drum	1	1'ID x 5' T-T	304 SS Clad	160	50
Acid Gas 1st Stg Compr KO Drum	1	1'ID x 3' T-T	304 SS Clad	160	100
Acid Gas 2nd Stg Compr KO Drum	1	1'ID x 3' T-T	304 SS Clad	160	230
Acid Gas Product KO Drum	1	1'ID x 3' T-T	304 SS Clad	160	670
Mol Sieve Dryer	2	6.5'ID x 17.5' T-T	304 SS Clad	450	670
Ammonia Separator	1	4'ID x 17' T-T	CS	200	235
Ammonia Surge Drum	1	12'ID x 34' T-T	CS	200	235
1st Stg Ammonia Compr KO Drum	1	8'ID x 26' T-T	CS	200	235
2nd Stg Ammonia Compr KO Drum	1	10'ID x 26' T-T	CS	200	235
Compressors					
Flash Gas Compressor	1	2,000 HP	CS	350	625
CO2 Sequestration Compressor	1	10,000 HP	CS	350	2400
Acid Gas Compressor	1	150 HP	CS	350	610
NH3 Refrig 1st Stage Compressor	1	1,600 HP	CS	200	136
NH3 Refrig 2nd Stage Compressor	1	20,000 HP	CS	350	245

Equipment	No. ea.	Size/Duty	Material	Temperature (°F)	Pressure (psig)
NH3 Refrig 3rd Stage Compressor	1	4,000 HP	CS	350	245
1st Stg Syngas Compressor	1	24,000 HP	CS	350	2600
1st Stg Syngas Expander	1	9,400 HP	CS	250	2400
Pumps					
Chilled Water Recycle Pumps (3-50% pumps)	3	3630 GPM/ 6,500 HP each	304 SS	200	1100
Cooling Water Pumps (3-50% pumps)	3	10,110 GPM/262 HP each	CS	200	100
Demineralized Water Pumps (2-100% pumps)	2	0.14 GPM/0.03 HP each	CS		
Lube Oil Pumps (3-50% pumps)	3	810 GPM/22 HP each	CS		

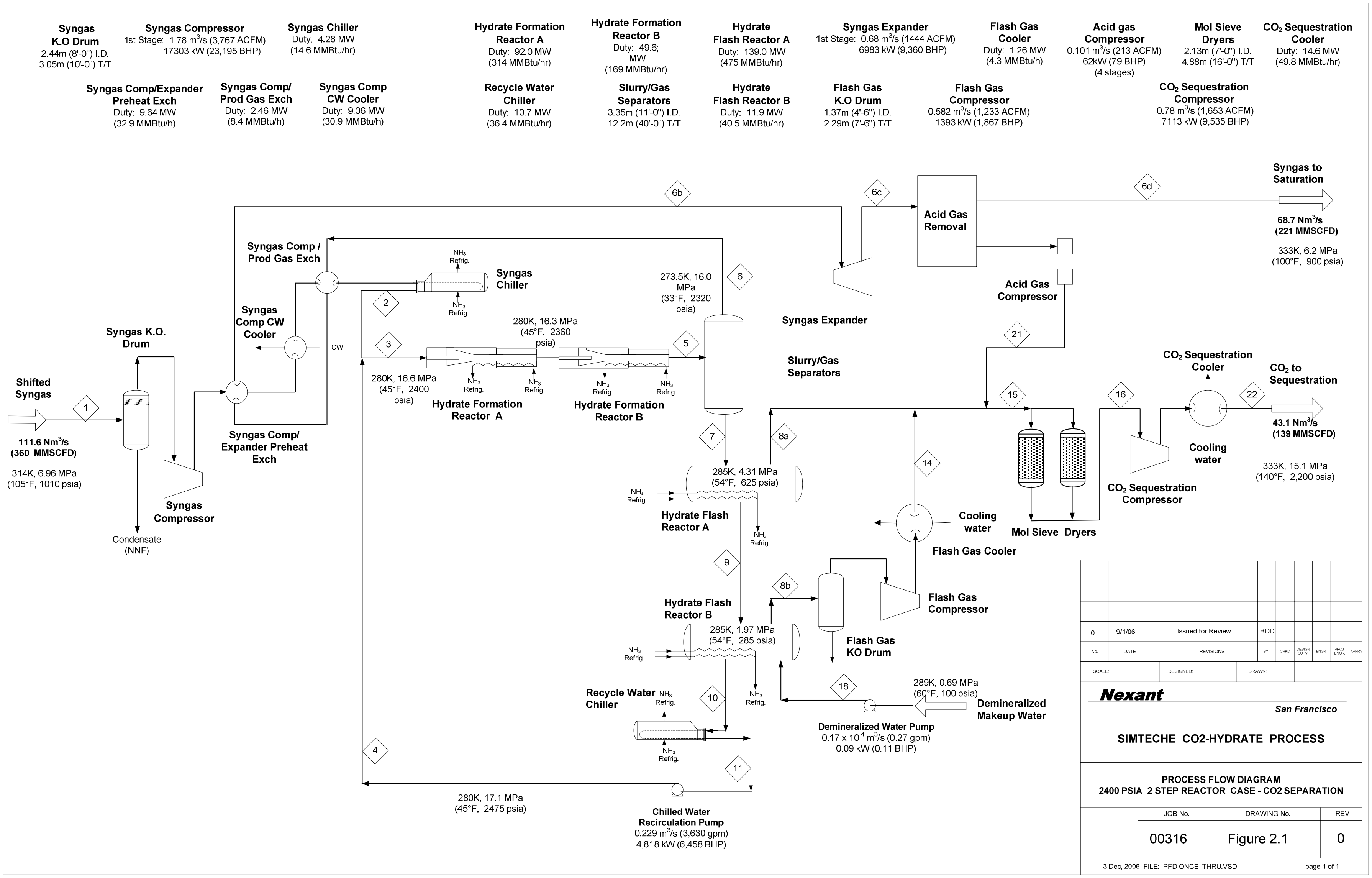
Table 1-7 Material Balance

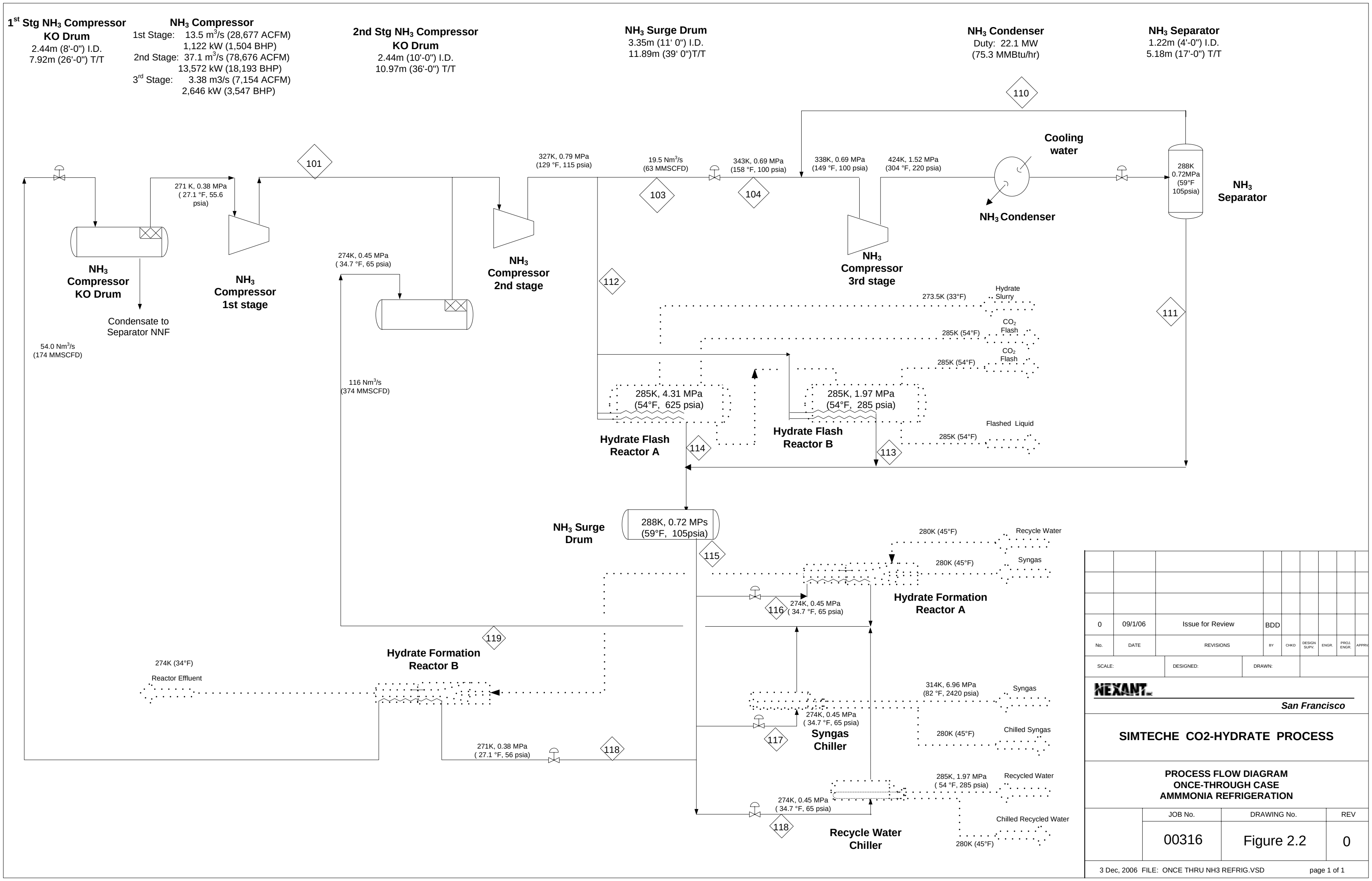
Stream Numbers		1	2	3	4	5A	5	6	6B	6C	6D	7
		Feed	Syngas Comp Outlet	Reactor Feed	Chill Water	Reactor A Effluent	Reactor Effluent	HP Sep Vapor	Expander Inlet	Expander Inlet	Syngas to Saturation	HP Sep Liquid
	Mol. Wt.	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr		lbmol/hr	lbmol/hr
H_2	2.02	21,418.72	21,418.72	21,418.76	0.04	21,418.76	21,418.76	21190.7145	21190.7145	21190.7145	21,190.71	228.05
CO_2	44.01	16,381.13	16,381.13	19,758.43	3377.3	9,158.20	3,178.30	1658.1	1658.1	1658.1	1,638.10	1520.2
H_2S	34.08	248.44	248.44	458.74	210.3	284.64	38.35	10.95	10.95	10.95	0.95	27.4
H_2O	18.02	0.00	0.00	0.00			0.97	0.97	0.96503736	0.96503736	0.95	
Ar	39.95	292.72	292.72	292.72		292.72	292.72	289.35	289.35372	289.35372	289.35	3.37
N2	28.01	289.32	289.32	289.32		289.32	289.32	285.99	285.99282	285.99282	285.99	3.33
CO	28.01	382.31	382.31	382.31		382.31	382.31	375.77	375.772499	375.772499	375.77	6.54
CH4	16.04	537.79	537.79	537.79		537.79	537.79	523.58	523.581588	523.581588	523.58	14.21
COS	60.07	0.04	0.04	0.06	0.02	0.06	0.06	0.02	0.01998	0.01998	0.02	0.04
NH3	17.03	3.4	3.40	384.40	381	384.40	384.40	1.24	1.23969	1.23969	1.24	383.16
H_2O (liquid)	18.02	-	0.00	197,538.97	197,538.97	132,892.99	95,534.88				0.00	95,534.88
CO_2 Hydrate	152.10	-	0.00	0.00		10,600.23	16,580.13				0.00	16,580.13
H_2S Hydrate	142.17	-	0.00	0.00		174.10	420.39				0.00	420.39
Total		39,553.87	39,553.87	241,061.50	201507.63	176,415.52	139,058.37	24,336.69	24,336.69	24,336.69	24,306.67	114,721.68

		lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr			lbs/hr	lbs/hr
H_2	2.02	43,176	43176	43176	0	43176	43176	42716	42716	42716	42716	460
CO_2	44.01	720,930	720930	869565	148634	403051	139876	72973	72973	72973	72092	66904
H_2S	34.08	8,466	8466	15632	7166	9699	1307	373	373	373	32	934
H_2O	18.02	-	0	0	0	0	17	17	17	17	17	0
Ar	39.95	11,694	11694	11694	0	11694	11694	11559	11559	11559	11559	134
N2	28.01	8,105	8105	8105	0	8105	8105	8012	8012	8012	8012	93
CO	28.01	10,709	10709	10709	0	10709	10709	10525	10525	10525	10525	183
CH4	16.04	8,628	8628	8628	0	8628	8628	8400	8400	8400	8400	228
COS	60.07	2	2	4	1	4	4	1	1	1	1	2
NH3	17.03	58	58	6546	6488	6546	6546	21	21	21	21	6525
H_2O (liquid)	18.02	-	0	3558704	3558704	2394094	1721080				0	1721080
CO_2 Hydrate	152.10	-				1612306	2521854				0	2521854
H_2S Hydrate	142.17	-				24751	59766				0	59766
Total		811,767	811,767	4,532,761	3,720,994	4,532,761	4,532,761	154,598	154,598	154,598	153,376	4,378,163
Temperature, oF		105	44.6	44.6	44.6	44.6	32.9	32.9	266	127	100	32.9
Pressure, Psia		1005	2400	2400	2475	2360	2320	2320	2320	922	900	2350

Table 1-7 Material Balance Cont'd

Stream Numbers		8A	8B	9	10	11	14	15	16	18	21	22
		Flash Rx A Vapor	Flash Rx B Vapor	625 Liquid	Recycle liquid	Chilled Water	Flash Gas	Total CO2 to Dryers	Total CO2 fr Dryers	Demin Water	Acid Gas AGR	Total CO2 to Sequest
	Mol. Wt.	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr							
H_2	2.02	225.30	2.71	2.75	0.04	0.04	2.71	228.01	228.01			228.01
CO_2	44.01	10,443.5	4279.5	7656.8	3377.3	3,377.30	4,279.50	14,743.03	14,743.03		20.00	14,743.03
H_2S	34.08	144.49	93	303.3	210.3	210.30	93.00	247.49	247.49		10.00	247.49
H_2O	18.02	3.58	3.17	0	0	0.00	3.17	8.26	0.00		1.52	0.00
Ar	39.95	3.33	0.04	0.04	0.00	0.00	0.04	3.37	3.37			3.37
N2	28.01	3.29	0.04	0.04	0.00	0.00	0.04	3.33	3.33			3.33
CO	28.01	6.43	0.11	0.11	0.00	0.00	0.11	6.54	6.54			6.54
CH4	16.04	13.86	0.35	0.35	0.00	0.00	0.35	14.21	14.21			14.21
COS	60.07	0.01	0.01	0.03	0.02	0.02	0.01	0.02	0.02			0.02
NH3	17.03	1.17	0.99	381.99	381.00	381.00	0.99	2.16	2.16			2.16
H_2O (liquid)	18.02			197,534.42	197,538.97	197,538.97	0.00	0.00	0.00	7.71		0.00
CO_2 Hydrate	152.10											
H_2S Hydrate	142.17											
Total		10,844.97	4,379.92	205,879.83	201,507.63	201,507.63	4,379.92	15,256.41	15,248.14	7.71	31.52	15,248.14
		lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr	lbs/hr
H_2	2.02	454	5	6	0	0	5	460	460		0	460
CO_2	44.01	459618	188340	336974	148634	148634	188340	648838	648838		880	648838
H_2S	34.08	4924	3169	10335	7166	7166	3169	8433	8433		341	8433
H_2O	18.02	64	57	0	0	0	57	149	0		27	0
Ar	39.95	133	2	2	0	0	2	134	134		0	134
N2	28.01	92	1	1	0	0	1	93	93		0	93
CO	28.01	180	3	3	0	0	3	183	183		0	183
CH4	16.04	222	6	6	0	0	6	228	228		0	228
COS	60.07	1	1	2	1	1	1	1	1		0	1
NH3	17.03	20	17	6505	6488	6488	17	37	37		0	37
H_2O (liquid)	18.02			3558622	3558704	3558704	0	0	0	139	0	0
CO_2 Hydrate	152.10	0										
H_2S Hydrate	142.17	0										
Total		465,708	191,600	3,912,456	3,720,994	3,720,994	191,600	658,556	658,408	139	1,248	658,408
Temperature, oF		54	54	54	54	44.6	105	68	67.3	60	242.7	140
Pressure, Psia		625	285	625	285	275	625	625	620	300	625	2200


Syngas K.O Drum
2.44m (8'-0") I.D.
3.05m (10'-0") T/T
Syngas Compressor
1st Stage: 1.78 m3/s (3,767 ACFM)
17303 kW (23,195 BHP)
Syngas Chiller
Duty: 4.28 MW
(14.6 MMBtu/hr)
Syngas Comp/Expander Preheat Exch
Duty: 9.64 MW
(32.9 MMBtu/h)
Syngas Comp/ Prod Gas Exch
Duty: 2.46 MW
(8.4 MMBtu/h)
Syngas Comp CW Cooler
Duty: 9.06 MW
(30.9 MMBtu/h)
Hydrate Formation Reactor A
Duty: 92.0 MW
(314 MMBtu/hr)
Hydrate Formation Reactor B
Duty: 49.6; MW
(169 MMBtu/hr)
Hydrate Flash Reactor A
Duty: 139.0 MW
(475 MMBtu/hr)
Syngas Expander
1st Stage: 0.68 m3/s (1444 ACFM)
6983 kW (9,360 BHP)
Flash Gas Cooler
Duty: 1.26 MW
(4.3 MMBtu/h)
Acid gas Compressor
0.101 m3/s (213 ACFM)
62kW (79 BHP)
(4 stages)
Mol Sieve Dryers
2.13m (7'-0") I.D.
4.88m (16'-0") T/T
CO2 Sequestration Cooler
Duty: 14.6 MW
(49.8 MMBtu/hr)
Recycle Water Chiller
Duty: 10.7 MW
(36.4 MMBtu/hr)
Slurry/Gas Separators
3.35m (11'-0") I.D.
12.2m (40'-0") T/T
Hydrate Flash Reactor B
Duty: 11.9 MW
(40.5 MMBtu/hr)
Flash Gas K.O Drum
1.37m (4'-6") I.D.
2.29m (7'-6") T/T
Flash Gas Compressor
0.582 m3/s (1,233 ACFM)
1393 kW (1,867 BHP)
CO2 Sequestration Compressor
0.78 m3/s (1,653 ACFM)
7113 kW (9,535 BHP)
Syngas to Saturation
68.7 Nm3/s
(221 MMSCFD)
333K, 6.2 MPa
(100°F, 900 psia)
6b
6c
6d
Acid Gas Removal
Syngas Comp / Prod Gas Exch
NH3 Refrig.
Syngas Chiller
273.5K, 16.0 MPa
(33°F, 2320 psia)
6
Acid Gas Compressor
Syngas Expander
Syngas Comp CW Cooler
CW
2
3
280K, 16.3 MPa
(45°F, 2360 psia)
5
Syngas K.O. Drum
Slurry/Gas Separators
21
CO2 Sequestration Cooler
CO2 to Sequestration
Shifted Syngas
1
280K, 16.6 MPa
(45°F, 2400 psia)
Hydrate Formation Reactor A
Hydrate Formation Reactor B
15
16
22
43.1 Nm3/s
(139 MMSCFD)
111.6 Nm3/s
(360 MMSCFD)
314K, 6.96 MPa
(105°F, 1010 psia)
Syngas Comp/ Expander Preheat Exch
7
8a
14
285K, 4.31 MPa
(54°F, 625 psia)
Cooling water
333K, 15.1 MPa
(140°F, 2,200 psia)
CO2 Sequestration Compressor
Syngas Compressor
Hydrate Flash Reactor A
Cooling water
Mol Sieve Dryers
Condensate (NNF)
Flash Gas Cooler
9
8b
Flash Gas Compressor
Hydrate Flash Reactor B
285K, 1.97 MPa
(54°F, 285 psia)
Flash Gas KO Drum
Recycle Water Chiller
10
18
289K, 0.69 MPa
(60°F, 100 psia)
Demineralized Makeup Water
Demineralized Water Pump
0.17 x 10-4 m3/s (0.27 gpm)
0.09 kW (0.11 BHP)
4
11
280K, 17.1 MPa
(45°F, 2475 psia)
Chilled Water Recirculation Pump
0.229 m3/s (3,630 gpm)
4,818 kW (6,458 BHP)
0 9/1/06 Issued for Review BDD
No. DATE REVISIONS BY CHKD DESIGN SUPV. ENGR. PROJ. ENGR. APPRV.
SCALE: DESIGNED: DRAWN:
Nexant
San Francisco
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
2400 PSIA 2 STEP REACTOR CASE - CO2 SEPARATION
JOB No. DRAWING No. REV
00316 Figure 2.1 0
3 Dec, 2006 FILE: PFD-ONCE_THRU.VSD page 1 of 1


1st Stg NH3 Compressor KO Drum
2.44m (8'-0") I.D.
7.92m (26'-0") T/T
NH3 Compressor
1st Stage: 13.5 m3/s (28,677 ACFM)
1,122 kW (1,504 BHP)
2nd Stage: 37.1 m3/s (78,676 ACFM)
13,572 kW (18,193 BHP)
3rd Stage: 3.38 m3/s (7,154 ACFM)
2,646 kW (3,547 BHP)
2nd Stg NH3 Compressor KO Drum
2.44m (10'-0") I.D.
10.97m (36'-0") T/T
NH3 Surge Drum
3.35m (11' 0") I.D.
11.89m (39' 0")T/T
NH3 Condenser
Duty: 22.1 MW
(75.3 MMBtu/hr)
NH3 Separator
1.22m (4'-0") I.D.
5.18m (17'-0") T/T
110
Cooling water
101
327K, 0.79 MPa
(129 °F, 115 psia)
19.5 Nm3/s
(63 MMSCFD)
343K, 0.69 MPa
(158 °F, 100 psia)
338K, 0.69 MPa
(149 °F, 100 psia)
424K, 1.52 MPa
(304 °F, 220 psia)
288K
0.72MPa
(59°F
105psia)
NH3 Separator
271 K, 0.38 MPa
(27.1 °F, 55.6 psia)
103
104
NH3 Condenser
274K, 0.45 MPa
(34.7 °F, 65 psia)
NH3 Compressor 2nd stage
NH3 Compressor 3rd stage
NH3 Compressor KO Drum
NH3 Compressor 1st stage
112
Hydrate Slurry
273.5K (33°F)
111
Condensate to Separator NNF
CO2 Flash
285K (54°F)
CO2 Flash
285K (54°F)
54.0 Nm3/s
(174 MMSCFD)
116 Nm3/s
(374 MMSCFD)
285K, 4.31 MPa
(54°F, 625 psia)
285K, 1.97 MPa
(54°F, 285 psia)
Flashed Liquid
285K (54°F)
Hydrate Flash Reactor A
114
Hydrate Flash Reactor B
113
NH3 Surge Drum
288K, 0.72 MPs
(59°F, 105psia)
115
280K (45°F)
Recycle Water
280K (45°F)
Syngas
274K, 0.45 MPa
(34.7 °F, 65 psia)
116
Hydrate Formation Reactor A
119
Hydrate Formation Reactor B
274K (34°F)
Reactor Effluent
314K, 6.96 MPa
(82 °F, 2420 psia)
Syngas
274K, 0.45 MPa
(34.7 °F, 65 psia)
Syngas Chiller
117
280K (45°F)
Chilled Syngas
271K, 0.38 MPa
(27.1 °F, 56 psia)
118
285K, 1.97 MPa
(54 °F, 285 psia)
Recycled Water
274K, 0.45 MPa
(34.7 °F, 65 psia)
118
Recycle Water Chiller
Chilled Recycled Water
280K (45°F)
0
09/1/06
Issue for Review
BDD
No.
DATE
REVISIONS
BY
CHKD
DESIGN SUPV
ENGR.
PROJ. ENGR.
APPRV.
SCALE:
DESIGNED:
DRAWN:
NEXANT INC.
San Francisco
SIMTECHE CO2-HYDRATE PROCESS
PROCESS FLOW DIAGRAM
ONCE-THROUGH CASE
AMMMONIA REFRIGERATION
JOB No.
DRAWING No.
REV
00316
Figure 2.2
0
3 Dec, 2006 FILE: ONCE THRU NH3 REFRIG.VSD
page 1 of 1